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# AUTOHEATED, AEROBIC, THERMOPHILIC DIGESTION OF MUNICIPAL SLUDGE WITH AIR AERATION

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16. ABSTRACT

A full-scale aerobic digestion system demonstrated that a simple self-aspirating aerator, that used ambient air, could achieve high oxygen transfer efficiencies and thereby allow conservation of heat. Continuous feed operation utilizing primary and waste activated sludges resulted in autoheated reactor temperatures ranging from 45 to 65 °C, even when ambient air temperatures were -20 °C and sludge temperatures were 0 °C.

Process variables included organic loading rate and D.O. concentration. It was observed that intermediate loading rates and low D.O. residuals (<1 ppm) allowed maximum temperature development.

The potential of the autoheated thermophilic digester to inactivate pathogens was investigated. Virus inactivation was 100%. Bacterial and parasite counts were less than those found in mesophilic anaerobic digestion sludge.

Dewaterability of the aerobically treated sludge deteriorated at all loading conditions studied. This last feature was the only disadvantage evident in the process.

17. KEY WORDS AND DOCUMENT ANALYSIS

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# FOREWORD

The U.S. Environmental Protection agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimonies to the deterioration of our natural environment. The complexity of that environment and the interplay of its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution; it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems to prevent, treat, and manage wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, to preserve and treat public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research and provides a most vital communications link between the researcher and the user community.

This full-scale study of the aerobic digestion process demonstrated that self-aspirating aerators could achieve high oxygen transfer and thus produce a well-stabilized, nearly pathogen-free sludge.

Francis T. Mayo, Director
Municipal Environmental Research
Laboratory

# ABSTRACT

Results are reported for a full-scale investigation of a simple sludge treatment system with the potential of providing a stabilized, pasteurized sewage sludge at lower costs than present aerobic systems. The goal of the 2-year study was to demonstrate that a simple autoheated aerobic digestion process using full-scale equipment could be used to treat sewage sludge.

A full-scale study (28.4 $m^3$ reactor) demonstrated that simple self-aspirating aerators (ambient air) could achieve high oxygen transfer efficiencies and thereby allow conservation of the heat of oxidation to achieve autoheating to high temperatures. Continuous-feed operation of a one-stage digestion system using a feed of primary and waste-activated sewage sludge (3% to 6% total solids) resulted in autoheated reactor temperatures ranging from 45°C to 65°C.

The relationship between process variables and the autoheating phenomena were examined at full-scale and bench-scale levels. The process variables included organic loading rate and dissolved oxygen concentration. Observation indicated that intermediate loading rates (12-15 kg TS/$m^3$ reactor-day and low dissolved oxygen residuals ($\leq$ 1 ppm) allowed maximum temperature development.

Two different aerators were tested and were found to achieve oxygen transfer efficiencies exceeding 20% at reactor temperatures that often exceeded 60°C. Operational problems associated with these kinds of aerators as well as with the other equipment on the thermophilic digestion facility were identified and examined.

The potential of the autoheated thermophilic digester to inactivate pathogens was investigated. Virus inactivation was 100% in most cases, with bacterial and parasite indicator counts less than those found in the full-scale mesophilic anaerobic digester effluent.

Dewaterability of the autoheated thermophilic digester effluent deteriorated at all loading conditions studied. The aerobic thermophilic digestion process appears to increase the solubility of various heavy metals such as cadmium.

A computer model was developed from the full-scale data for predicting the reactor temperature under given loading conditions.

This report was submitted in fulfillment of Grant Number R804636-01 by Cornell University under the sponsorship of the U. S. Environmental Protection Agency. This report covers the period October 13, 1976 to December 1, 1978. Work was completed October 12, 1979.

CONTENTS

Abstract. . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . iv
Figures . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . viii
Tables. . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . . xv
Abbreviations and Symbols . . . . . . . . . . . . . . . . . . . . . . . xviii
Acknowledgments . . . . . . . . . . . . . . . . . . . . . . . . . . . . xxi

1. Introduction . . . . . . . . . . . . . . . . . . . . . . . . . . 1
2. Conclusions. . . . . . . . . . . . . . . . . . . . . . . . . . . 5
3. Recommendations. . . . . . . . . . . . . . . . . . . . . . . . . 7
4. Background and Theory. . . . . . . . . . . . . . . . . . . . . . 9
5. Investigative Facilities, Binghamton-Johnson City Sewage Treatment Plant Operation, and System Design and Construction . . . . . . . . . . . . . . . . . . . . . . . 28
6. Materials and Methods. . . . . . . . . . . . . . . . . . . . . . 67
7. Results. . . . . . . . . . . . . . . . . . . . . . . . . . . . . 84
8. Discussion of Experimental Results . . . . . . . . . . . . . . . 182
9. Engineering Significance and Economic Considerations . . . . . . 191

References. . . . . . . . . . . . . . . . . . . . . . . . . . . . . . 202
Appendices. . . . . . . . . . . . . . . . . . . . . . . . . . . . . . 210

Preceding page blank

# FIGURES

Number Page

1 Computer model prediction of heat energy losses in aerobic digestion of sewage sludge (2) . . . . . . . . . . . . . . . . 11

2 Comparison of the autoheating potential using air aeration and pure oxygen in sewage sludge treatment (2) . . . . . . . 12

3 Aerator performance of the thermophilic digestions facility's self-aspirating aerator when treating cow manure diluted to 5% solids. OLR is the biodegradable COD loading rate (26). . . . . . . . . . . . . . . . . . . . . . . . . 16

4 Relationship of loading rate, solids retention time, substrate removal rate, and substrate removal efficiencies (26). . . . . . . . . . . . . . . . . . . . . 17

5 Relationships between energy and organic conversion efficiencies . . . . . . . . . . . . . . . . . . . . . . . . 22

6 Illustration of parameters considered in the energy balance model of a completely mixed biological reactor (26). . . . . 24

7 An aerial view diagram illustrating the Binghamton-Johnson City STP components and relative location of Cornell University's digestion facility. . . . . . . . . . . . . . . 29

8 Floor plan and analytical laboratory in the Cornell Agricultural Waste Management Laboratory . . . . . . . . . . 30

9 Influent wastewater flow variations during the period of full scale study . . . . . . . . . . . . . . . . . . . . . . 34

10 $BOD_5$ data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978 . . . . . . . . . . . . . . . . . . 36

11 TSS data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978 . . . . . . . . . . . . . . . . . . 37

12 TCOD data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978 . . . . . . . . . . . . . . . . . . 38

Number Page

13 Volumetric loading rate, kg $BOD_5/m^3$-day, for the Binghamton-Johnson City aeration basins during the experimental test period . . . . . . . . . . . . . . . . . . 39

14 Schematic diagram of the Binghamton-Johnson City STP major unit processes and their respective average performances . . 41

15 Sludge flows for raw, return, and excess sludges during the experimental test period . . . . . . . . . . . . . . . . . 42

16 MLSS and return sludge SS concentrations of the Binghamton-Johnson City STP for the period of the full scale study. . . 44

17 Settling property, as measured by SVI, observed for the activated sludge mixed liquor during the experimental test period. . . . . . . . . . . . . . . . . . . . . . . . . 45

18 Concentration of TS from the thickener cone(s) during the experimental test period . . . . . . . . . . . . . . . . 47

19 Site of the thermophilic digester test unit location at the Binghamton-Johnson City STP. The unit in the background is one of the sludge thickeners from which sludge was obtained for the study . . . . . . . . . . . . . . . . . . 50

20 Site of the thermophilic digester study at the Binghamton-Johnson City STP showing the effluent storage tank prior to installation. . . . . . . . . . . . . . . . . . . . . . . . 50

21 Thermophilic digestion study effluent tank during installation at the Binghamton-Johnson City STP. . . . . . . 51

22 Thermophilic digestion system at the Binghamton-Johnson City STP prior to insulation and piping additions. . . . . . 51

23 Setting the reactor tank for the thermophilic digestion facility at the Binghamton-Johnson City STP. . . . . . . . . 52

24 Electrical control panel fabricated for the thermophilic digestion facility . . . . . . . . . . . . . . . . . . . . 52

25 Aerobic thermophilic test facility . . . . . . . . . . . . . . 53

26 Aerobic thermophilic test facility located beside the sludge thickeners at the Binghamton-Johnson City STP. . . . . . . . 53

27 Side view of the completed 28.4 $m^3$ insulated thermophilic reactor, feed tank, and effluent tank. . . . . . . . . . . . 54

Number Page

28 Relationship of the thermophilic digestion test facility to other unit processes at the Binghamton-Johnson City STP. . . . . . . . . . . . . . . . . . . . . . . . 54

29 Completed thermophilic digestion system as it appeared in the second year of the study . . . . . . . . . . . . . . 55

30 Binghamton-Johnson City STP full scale thermophilic digestion facility. . . . . . . . . . . . . . . . . . . . . . 56

31 Aerial view of Binghamton-Johnson City full scale thermophilic digestion facility, showing major components. . . . . . . . . . . . . . . . . . . . . . . . . 57

32 Comparison of the two aerators tested in this study and their location in the reactor . . . . . . . . . . . . . . 59

33 The centri-rator aerator suspended above the reactor (top photograph), and in place in the reactor (bottom photograph) . . . . . . . . . . . . . . . . . . . 60

34 The Midland-Frings aerator on top of the reactor without support tripod and air intake tube in place . . . . . 61

35 Continuously and semi-continuously fed laboratory scale reactors. . . . . . . . . . . . . . . . . . . . . . . 65

36 Batch-mode laboratory scale reactor systems . . . . . . . . 66

37 Schematic diagram showing the reactor with aeration testing equipment in place. . . . . . . . . . . . . . . . . . 69

38 Graphical determination of the refractory fraction of a substrate. . . . . . . . . . . . . . . . . . . . . . . . 75

39 Graphical determination of the refractory fraction of a substrate by reciprocal plot. . . . . . . . . . . . . . . 75

40 Typical cooling in the full scale aerobic digester. . . . . 88

41 Static and dynamic heat losses as related to the average temperature difference (reactor and ambient) for full scale, non-steady state, tap water tests. . . . . . . . . 90

42 TCOD removal as related to time of operation, biodegradability Study No. 4, Unit No. 1 . . . . . . . . . . . 93

43 TS removal as related to time of operation, biodegradability Study No. 4, Unit No. 1 . . . . . . . . . . . 94

| Number | | Page |
|---|---|---|
| 44 | TVS removal as related to time of operation, biodegradability Study No. 4, Unit No. 1 . . . . . . . . . . . | 95 |
| 45 | TKN and NH$_3$-N removal as related to time of operation, biodegradability Study No. 4, Unit No. 1. . . . . . . . . | 96 |
| 46 | Change in sludge dewaterability (CST) as related to time of operation, biodegradability Study No. 4, Unit No. 1. . . . . . . . . . . . . . . . . . . . . . . | 97 |
| 47 | Change in sludge dewaterability, CST and ZSR, as related to time of operation and TVS destruction. . . . . . . . . | 99 |
| 48 | Relationship of pH with time of operation, biodegradability Study No. 4, Unit No. 1 . . . . . . . . . . . . . . | 100 |
| 49 | Rate of autoheating during start up of aerobic sludge digester during summer and winter conditions. . . . . . . | 102 |
| 50 | Reactor (autoheated), ambient, and feed temperatures with corresponding loading rates and operational modes for the duration of the full scale study (May 18, 1977 - November 24, 1978) . . . . . . . . . . . . . . . . . | 107 |
| 51 | Relationship observed for removal rate (BCOD) versus loading rate (BCOD) and hydraulic retention time (HRT)$^+$ for bench and full scale systems at steady state conditions. ($^+$HRT based on constant feed concentration of 40 gm BCOD/ℓ.) . . . . . . . . . . . . . . | 118 |
| 52 | Relationship observed for removal efficiency (BCOD) versus loading rate (BCOD) and hydraulic retention time (HRT)$^+$ for bench and full scale systems at steady state conditions ($^+$HRT based on constant feed concentration of 40 gm BCOD/ℓ.). . . . . . . . . . . | 119 |
| 53 | Reaction rate coefficient as related to organic loading rate (BCOD) for both bench and full scale studies (HRT based on constant feed concentration of 40 gm BCOD/ℓ.). . . . . . . . . . . . . . . . . . . . . . . . . | 120 |
| 54 | Reactor rate coefficient ($K_T$) as a function of temperature (T) for full scale and bench scale reactors . | 121 |
| 55 | Exponential relationship observed for the reaction rate coefficient ($K_T$) as a function of temperature . . . . . . | 122 |

Number Page

56 Effect of temperature on the reaction rate coefficient ($K_T$) (compared to $K_T$ at 20°C) when $K_{T\,20} = 0.10$ day$^{-1}$ and biodegradable substrate $\cong$ 75% of total. . . . . . . . 123

57 Temperature differential, between the autoheated reactor and ambient temperature observed at steady conditions, at various biodegradable organic loading rates. . . . . . 125

58 Temperature differential, between the autoheated reactor and feed sludge temperature, observed at steady state conditions at various biodegradable organic loading rates . . . . . . . . . . . . . . . . . . . . . . . . . 126

59 Steady state heat balance parameters as a function of BCOD loading for the Midland-Frings aerator . . . . . . . 127

60 Steady state heat balance parameters as a function of BCOD loading for the centri-rator aerator . . . . . . . . 128

61 Illustration of organic removal rate (BCOD) and total biological heat production (BHP) as related to organic loading rate for the full scale system at steady state conditions. . . . . . . . . . . . . . . . . . . . . . 130

62 Quantity of heat released during microbial respiration as a function of BCOD loading for both the centri-rator and Midland-Frings aerators . . . . . . . . . . . . 132

63 Observed relationship between pH and solids loading rates for full scale study. . . . . . . . . . . . . . . . . . . . 132

64 Distribution profiles of dissolved oxygen concentrations with depth within the reactor at various organic loading rates . . . . . . . . . . . . . . . . . . . . . . 134

65 Distribution of dissolved oxygen concentrations with depth within the reactor after the addition of a concrete radius . . . . . . . . . . . . . . . . . . . . . 136

66 Oxygen transfer efficiency, as measured by TCOD removal-rate analysis, as related to organic loading rate . . . . 138

67 Oxygen transfer rate (O.T.R.) as related to organic loading rate (BCOD) and hydraulic retention time. (HRT based on constant feed concentration of 40 gm BCOD/ℓ.). . . . . 139

68 Airflow as related to organic loading (BCOD) rate for the centri-rator. . . . . . . . . . . . . . . . . . . . . 140

Number Page

69 Relationship of bubble diameter to the area available for gas transfer. . . . . . . . . . . . . . . . . . . . . . 141

70 Photograph illustrating the bubble photography apparatus. . 144

71 Aeration bubbles as created by self-aspirating aerators operating in sewage sludge at thermophilic temperatures . 145

72 Foam bubbles characteristic of the foam layer generated during autoheated aerobic digestion of sewage sludge. . . 146

73 *Salmonella* sp. as found in raw, anaerobic, and aerobic sludges ($\leq$ 3 organisms/100 ml considered below limits of detection) . . . . . . . . . . . . . . . . . . . . . . 152

74 Results of analyses for *Pseudomonas aeruginosa* in raw, anaerobic, and aerobic sludges . . . . . . . . . . . 153

75 Results of virus analyses (total PFU/100 ml wet sludge) on raw, anaerobically (mesophilic), and aerobically (thermophilic) digested sludge. . . . . . . . . . . . . 157

76 Total parasite ova found in raw, anaerobic, and aerobic sludge for all dates of collection as analyzed by Tulane University . . . . . . . . . . . . . . . . . . . . . 161

77 Cadmium solubilization from raw, thermophilic aerobic, and mesophilic anaerobic sludges acidified to a pH of 2.0. . . . . . . . . . . . . . . . . . . . . . . . . 163

78 The influence of biodegradable substrate in the influent sludge on the predicted reactor temperature as given by the computer model . . . . . . . . . . . . . . . . . . 168

79 The influence of biodegradable organics (BCOD) on reactor temperature at various retention times (HRT) assuming $K_T = 0.4$ day$^{-1}$ and R = 3.5 Kcal/gm with curves generated by computer model . . . . . . . . . . . . . . . . . . . . . 169

80 The reactor temperature shown as both a function of biodegradable loading rate and reaction rate as predicted by the computer model . . . . . . . . . . . . . . . . . . 170

81 The relationship between treatment efficiency and loading rate for various values of $K_T$ as predicted by the computer model with all real steady state (average) values also shown . . . . . . . . . . . . . . . . . . . . . 171

Number Page

82 Diagram of the centrifugal feed pump illustrating the various components. . . . . . . . . . . . . . . . . . . . 176

83 Photograph of centri-rator impeller . . . . . . . . . . . . 177

84 Diagram of centri-rator aerator . . . . . . . . . . . . . . . 178

85 Centri-rator styrofoam float system illustrating deterioration of the styrofoam blocks . . . . . . . . . . 179

86 Midland-Frings aerator with accumulated rag debris in sludge intake area. . . . . . . . . . . . . . . . . . . . . 181

87 Thermophilic aerobic digestion system . . . . . . . . . . . 192

88 Influence of single and series multiple stage, completely-mixed reactors on substrate removal efficiencies with two different substrate removal rate coefficients. Each series reactor has a one-day HRT . . . . . . . . . . 195

89 Influence of single and series multiple stage, completely-mixed reactors on substrate removal efficiency with two different substrate removal rate coefficients. Each series reactor system has three stages with one-third the volume (per stage) of a one-stage system. . . . 196

90 Comparison of capital, operation, and maintenance cost(s) for sludge treatment. . . . . . . . . . . . . . . . . . . . 200

# TABLES

| Number | | Page |
|---|---|---|
| 1 | Previous large scale experiences with autoheated aerobic digestion of sewage sludges . . . . . . . . . . | 14 |
| 2 | Pathogenic organisms associated with wastewater sludges . | 19 |
| 3 | Free energy of substrate oxidation as related to the theoretical COD for aerobic oxidations (71) . . . . . . | 21 |
| 4 | Design specifications of major process units. . . . . . . | 32 |
| 5 | Wastewater flows for the Binghamton-Johnson City STP (74) | 33 |
| 6 | Source and composition of wastewater discharged to the Binghamton-Johnson City STP (74). . . . . . . . . . . . | 35 |
| 7 | Binghamton-Johnson City STP performance summary for major unit processes. . . . . . . . . . . . . . . . . . | 40 |
| 8 | Detailed summary of average operational parameters for the activated sludge unit(s) at the Binghamton-Johnson City STP. . . . . . . . . . . . . . . . . . . . . . | 43 |
| 9 | Summary of full scale construction in order of succession. . . . . . . . . . . . . . . . . . . . . . . | 49 |
| 10 | Kinds of analyses and frequency of testing. . . . . . . . | 77 |
| 11 | Major project activities. . . . . . . . . . . . . . . . . . | 85 |
| 12 | Results of non-steady state clean water testing of centri-rator aeration system. . . . . . . . . . . . . . . | 86 |
| 13 | Determination of the heat loss characteristics of the reactor during aeration of tap water. . . . . . . . . . | 89 |
| 14 | Determination of the heat loss characteristics of the reactor during quiescent conditions . . . . . . . . . . | 89 |
| 15 | Results of seven biodegradability studies covering the period August 17, 1977 to August 29, 1978 . . . . . . . | 98 |

Number Page

16 Performance data as averaged for steady state conditions for both full and bench scale digesters . . . . . 104

17 Summary of steady state heat balance and kinetic parameters for bench and full scale systems . . . . . . . 129

18 Summary of oxygen transfer results for steady state full scale operation. . . . . . . . . . . . . . . . . . . 137

19 Pathogen analyses conducted on samples of raw primary and waste activated sludge mixtures, and effluent from the mesophilic anaerobic digester and the thermophilic aerobic digester. . . . . . . . . . . . . . . . . . . . . . 147

20 Operational data for dates of pathogen sampling . . . . . . 148

21 Results of bacterial analyses for all test dates. All analyses conducted by the U.S. Environmental Protection Agency, Cincinnati, Ohio. . . . . . . . . . . . . 150

22 Summary of the fate of fecal coliforms and fecal streptococci during mesophilic anaerobic digestion and thermophilic aerobic digestion of primary and waste activated sludge mixtures . . . . . . . . . . . . . . . . 151

23 Results of qualitative analysis for enteric pathogens by Cornell University Veterinary Diagnostic Laboratory (100 ml wet sludge analyzed). . . . . . . . . . . . . . . 154

24 Results of virus analyses for total PFU of primary and waste activated sludge mixtures. Analyses conducted by the U.S. Environmental Protection Agency Laboratory . . . 155

25 Results of *Ascaris* analyses in primary and waste activated sludge. Analyses conducted by the University of Cincinnati. . . . . . . . . . . . . . . . . . . . . . . 158

26 Summary of parasite analyses for all dates of collection for primary and waste activated sludge combinations. All analyses conducted by the Department of Tropical Medicine, Tulane University . . . . . . . . . . . . . . . 159

27 Results of parasitologic examinations of all samples of primary and waste activated sludge combinations. All analyses conducted by Tulane University . . . . . . . . . 160

28 Percentage of heavy metals solubilized in raw, mesophilic anaerobic, and thermophilic aerobic sludges acidified with nitric acid (pH = 2.0) for twelve hours. . . . . . . 164

Number | Page

29 Computer model program for prediction of auto-heated reactor temperature. . . . . . . . . . . . . . . . . . . 166

30 Computer program printout example, where symbols are as defined earlier. . . . . . . . . . . . . . . . . . . . . 167

31 Comparison of steady state performance with computer model results . . . . . . . . . . . . . . . . . . . . . . . 172

32 Problems experienced with the full scale system and the associated dates. . . . . . . . . . . . . . . . . . . . . 173

33 Comparison of design specifications and reported cost for sewage sludge treatment. . . . . . . . . . . . . 199

## LIST OF ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

| | |
|---|---|
| A | --area |
| a | --intercept coefficient |
| ASTARC | --Animal Science Teaching and Research Center (Cornell U.) |
| AWML | --Agricultural Waste Management Laboratory (Cornell U.) |
| b | --slope coefficient |
| BCOD | --biodegradable chemical oxygen demand |
| BHP | --biological heat production |
| BOD | --biochemical oxygen demand |
| $C_d$ | --coefficient of diffusion |
| Cd | --cadmium |
| $C_{PL}$ | --specific heat capacity |
| $C^*_s$ | --concentration of dissolved oxygen at given conditions |
| $C_t$ | --concentration of dissolved oxygen at time, t |
| CST | --capillary suction time |
| D | --density |
| D.O. | --dissolved oxygen concerntration |
| e | --subscript denoting effluent |
| F/M | --food to microorganism ratio |
| g | --force of gravity |
| $\gamma$ | --specific weight |
| $H_A$ | --heat loss to the surface area of the reactor $H_E$ |
| $H_E$ | --heat loss to the effluent |
| $H_L$ | --overall heat loss coefficient |
| $H_{LA}$ | --heat loss through aeration |
| $H_M$ | --heat input due to mixing |
| $H_S$ | --heat loss as sensible heat |
| $H_V$ | --heat loss as latent heat of vaporization |
| HRT | --hydraulic retention time |

hp --horsepower
K --reaction rate coefficient
KW --kilowatt
$K_{LA}$ --oxygen transfer coefficient
MGD --million gallons per day
ML --mixed liquor
N --oxygenation capacity
$NH_3$-N --ammonia nitrogen
o --subscript denoting influent
OLR --organic loading rate
ORP --oxidation reduction potential
OTE --oxygen transfer efficiency
OTR --oxygen transfer rate
PFU --plaque forming units (virus)
$PT_R$ --predicted reactor temperature
Q --flow
r --subscript denoting return flow or concentration
R --heat released during microbial respiration (as used in steady state performance calculations)
R --resistance to heat transfer (as used to describe the insulation values of materials)
R --refractory fraction of an organic parameter (as used in biodegradability determinations)
$r^2$ --coefficient of determination
RNA --ribonucleic acid
rpm --revolutions per minute
RR --removal rate
S --substrate concentration
SOR --surface overflow rate
SRT --solids retention time
STP --sewage treatment plant
SVI --sludge volume index
t --time
T --subscript denoting first order reaction rate coefficient

| | |
|---|---|
| $T_A$ | --ambient temperature |
| TE | --treatment efficiency |
| TKN | --total Kjeldahl nitrogen |
| $T_o$ | --influent substrate temperature |
| $T_R$ | --reactor temperature |
| TS | --total solids |
| TSS | --total suspended solids |
| TVS | --total volatile solids |
| TVSS | --total volatile suspended solids |
| $\Theta$ | --temperature coefficient (Van't Hoff-Arrhenius) |
| $\Theta_c$ | --mean cell residence time |
| $\Theta_h$ | --hydraulic residence time |
| $\mu$ | --micron |
| V | --volume |
| V | --velocity |
| w | --subscript denoting waste flow or concentration |
| X | --microorganism concentration |
| Z | --elevation |
| ZSR | --zone settling rate |

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## SECTION 1

## INTRODUCTION

Sewage sludge and other waste organics should be used for beneficial purposes whenever such practices would be safe and cost effective. Implementation of this policy usually results in the application of sludge to agricultural land for soil property improvement or fertilization for crop production. If the land application rates of sewage sludge were limited to the plant nitrogen requirements, sewage sludge would provide the required plant nutrients for millions of acres of cropland. However, present sludge management technology often cannot guarantee public health protection or cost effective solutions. This report summarizes the results of a full scale investigation of a simple sludge treatment system that has the potential of providing a stabilized, pasteurized sewage sludge at costs less than present aerobic processing systems.

Approximately 20 percent of the sewage sludges produced in the United States are used in agriculture and this practice is expected to increase as ocean dumping and other disposal practices are terminated (1). Of course, alternative methods are known that stabilize and pasteurize sewage sludge, but few if any can achieve the high level of treatment required at a low cost without significant disadvantages.

Among the options available for a more effective sludge treatment approach is thermophilic biological treatment. Both anaerobic and aerobic treatment processes can operate in the thermophilic temperature zone of 43°C to 70°C. Autoheated aerobic digestion discussed here refers to a process which uses the metabolic heat of oxidation of the organics to increase the temperature of the aerating slurry.

The theoretical heat energy required to increase sludge temperatures to the thermophilic range would be about 40 Kcal per liter and at today's energy prices would cost about $10 per million gallons of sewage flow, or about $20 per ton of dry sludge. However, the heat generated in composting of organics by microbial thermogenesis is sufficient to achieve high temperatures or drive what has been referred to as autothermal processes (2). In fact, aerobic industrial fermentations generate so much heat that cooling methods are an integral part of the design of these reactors in biochemical engineering. This study attempted to demonstrate that the heat of oxidation of sewage sludge could be used to achieve high temperature aerobic treatment of sewage sludge.

Both anaerobic and aerobic treatment processes can operate at the thermophilic zone (43°C to 70°C). In order to avoid adding the external heat to the reactor, it would be ideal if the process could supply most of the required energy. The following simplistic comparison of the energy needs to operate aerobic and anaerobic systems at high temperatures can be used to illustrate the feasibility of this alternative. The minimum energy needs are assumed to be equal to the amount of energy needed to heat the sludge to thermophilic temperatures (greater than 43$^{\circ}$C). If it is assumed that in the worst case the influent temperature of the sludge is 10$^{\circ}$C and the desired operating temperature is 60$^{\circ}$C, the minimum required energy would be that energy required to raise the liquid temperature 50$^{\circ}$C, or about 50 Kcal/liter. If it is assumed that the sludge, in this example, has a 2 percent volatile solids concentration and that 50 percent of the organics are stabilized in the reactor, this would result in production of methane gas with an energy value of 50 Kcal/liter of sludge processed. Since conversion of this biogas to hot water is only about 75 percent efficient, the maximum reactor temperature that would be achieved would be 47.5$^{\circ}$C using methane produced by anaerobic digestion, assuming zero heat losses from the system.

The aerobic process has one potential source of energy for autoheating, i.e., the heat released to the environment during aerobic oxidation of the substrate. It is knownb that this averages about 3.5 Kcal of heat released for every gram of chemical oxygen demand (COD) oxidized. Thus, if the above sludge is used at 2 percent volatile solids, the total amount of heat generated would be 50 Kcal/liter Theoretically, this would provide the energy necessary to heat the cold sludge to 60$^{\circ}$C, assuming no heat losses. Thus, it would appear that aerobic treatment would have potential in this area. Therefore, the challenge is to identify means of managing the heat of microbial oxidation in liquid slurries in order to maximize the autoheating capabilities.

The potential advantages of high-temperature aerobic sludge treatment are thought to include:

-increased rates of oxidation, thus resulting in smaller digester volume requirements;
-increased quantity of organics stabilized;
-destruction of most pathogenic bacteria, virus, and parasites;
-significantly smaller oxygen requirements as a result of elimination of nitrification over ambient temperature aerobic digestion;
-increased ease of liquid-solid separation; and
-weed seeds destruction.

If these advantages are applied to sludge treatment with little or no cost increase over conventional processes, it would represent significant improvement in sludge treatment technology.

## PROBLEM MAGNITUDE

One of the objectives of the Water pollution Control Act Amendments in Pub. L. No. 92-500 is to provide all domestic wastewaters with a minimum of secondary treatment. For every 10,000 people, about 38 m$^3$ per day (10,000 gallons per day) of sewage sludge will be generated (at about 4% dry solids concentration) with the application of secondary treatment. When this objective is achieved, the quantity of domestic sewage sludge which must be disposed of will be greater than 0.8 million cubic meters per day. If this material is to be distributed on private land, used for food production, or used indiscriminately by homeowners, it is necessary that the public health aspects of the material be a prime concern.

Long term anaerobic digestion at 35°C has been reported to effectively control the majority of human pathogens, but some may survive for long periods (3). In smaller, less well operated sewage facilities, the sludge may not receive as effective treatment as in the larger plants with efficient anaerobic digesters. Also, because of the complexities of the anaerobic digestion process, many small town sewage facilities use the simpler aerobic digestion process. It is known that many groups of pathogens can survive the ambient temperature treatment received in many small aerobic digestion facilities (4).

Although the possibilities of the existence of pathogens in treated sewage sludge is fairly well known, this has not inhibited some individuals from using this material in agriculture. In an informal survey in 1965 at the University of Illinois (5), it was found that over 300 communities were selling the local sewage sludge by advertising in the newspaper, or were making it available to anyone who wanted to use it on lawns and gardens. Bastian (1) has estimated that about 20 percent of all sewage sludge is presently used in agriculture.

Thus, it is clear that a simple process that would be capable of providing more effective sludge treatment and pathogen control at a cost not exceeding existing sludge treatment cost should have a high priority in new technology development. This study attempted to determine the prospect(s) of adapting existing technology and knowledge to produce a new simple process capable of yielding a pasteurized, stabilized sludge, without the use of a large amount of energy.

## GOALS AND OBJECTIVES

The general goal of this two-year study was to demonstrate that a simple autoheated aerobic digestion process using full scale equipment that has been shown to be effective with animal wastes (6,7) could be used to treat sewage sludge. Specific objectives of this study were to:

(1) demonstrate the feasibility of achieving autoheating to temperatures exceeding 43°C with typical primary and secondary waste activated sludge using a simple air aeration system;

(2) estimate the practical operating problems of a full scale system over an extended period of operation;

(3) develop the relationships between the sludge autoheating characteristics and the volumetric and organic loading rates to enable prediction of the reactor temperature;

(4) define the aeration requirements and heat balance needed to achieve autoheating with air aeration;

(5) measure the high temperature treatment impact on dewaterability of the sludge; and to

(6) determine the effectiveness of the autoheated aerobic system on the destruction of major groups of pathogens and compare these results to the existing anaerobic digestion system at the sewage treatment facility.

This report is the summary of the two-year project. Results of this study were also reported elsewhere (8,9,10).

# SECTION 2

# CONCLUSIONS

1. This two-year demonstration focussed on operating a full scale system for sludge treatment for populations exceeding 5,000 people. Mixtures of thickened waste activated sludge and primary sludge were autoheated to temperatures in excess of 43°C for all conditions tested using self-aspirating air aerators and well insulated tanks in a full scale reactor volume of 28.4 $m^3$ during 1.5 years of operation. Autoheated slurry temperatures normally exceeded 50°C and reached a maximum of 65°C.

2. These results indicate that a simple aerobic digestion system can achieve autoheating to thermophilic temperatures with typical sewage sludges (50% biodegradable TVS or greater) at concentrations greater than 2 percent total solids under cold weather conditions. Use of heat exchangers would enable even more dilute sludges to be autoheated to high temperatures.

3. Thermophilic aerobic reactors are easily started and will achieve temperatures in excess of 40°C in approximately 10 days when sludge temperatures are as low as 0°C and air temperatures are minus 10°C.

4. Maximum autoheated temperatures were obtained at loading rates of between 3 and 10 kg of biodegradable COD per $m^3$ per day with the maximum organic removal rate (6.5 kg BCOD/$m^3$-day) obtained at a loading rate of 8 kg BCOD/$m^3$-day for the Midland-Frings aerator. Approximately 75 percent of the biodegradable organics were oxidized at this loading rate. This corresponded to a 5-day hydraulic retention time with the sludges used in this study at total solids concentrations of approximately 5 percent.

5. The efficient conservation of the heat of organic oxidation was achieved using two kinds of self-aspirating aerators that were shown to be capable of achieving greater than 20 percent oxygen transfer efficiency. At high loading rates, oxygen transfer efficiencies exceeded 23 percent with air aeration for both aerators tested with organic slurries up to 5 percent total solids at temperatures exceeding 50°C. Normal oxygen transfer analysis using comparative tap water testing does not apply to these kinds of aeration systems.

6. The dewaterability of the treated sludge deteriorated significantly at all HRTs tested (3 to 11 days).

7. The conservation of nitrogen through the system was observed at all conditions tested and was a function of high temperature inhibition of nitrification and minimal volatilization of ammonia.

8. The thermophilic aerobic digester exhibited a high degree of pathogen control with respect to bacteria, viruses, and parasites. Complete inactivation, i.e., below the limits of detection, of Salmonella sp. and total plaque

forming units (viruses) was observed for all but one test date. Parasite numbers (viable ova) were significantly reduced by the aerobic thermophilic system, but not completely.

9. Heavy metals, e.g., cadmium, appear to become more soluble with thermophilic aerobic digestion.

10. Practical operational and maintenance problems were found to be minimal and a function of the temporary nature of the digestion facility.

11. Daily operational requirements of the aerobic thermophilic system are minimal.

12. Major requirements for autoheated aerobic digestion of sewage sludge include a minimum biodegradable organic concentration, an insulated reactor, and the use of efficient aerators such as the two tested in this study. The potential for autoheating can be predicted using energy and mass balances. The following equation was developed for the prediction of the autoheated reactor:

$$PT_R = \frac{HRT\ K_T\ BCOD_e\ R}{D_L\ C_{PL}} - [0.0407\ (T_R - T_A)^{1.2765}]\ HRT + T_O$$

where $PT_R$ = predicted reactor temperature, °C
$HRT$ = hydraulic retention time, day
$K_T$ = reaction rate coefficient, $\text{day}^{-1}$
$BCOD_e$ = biodegradable COD in the effluent, gm/ℓ
$R$ = heat released during microbial respiration, Kcal/gm COD
$D_L$ = density of the liquid sludge, gm/ℓ
$C_{PL}$ = specific heat capacity of the liquid sludge, cal/gm-°C
$T_R$ = reactor temperature, °C
$T_A$ = ambient temperature, °C
$T_O$ = feed sludge, temperature, °C.

13. The best practical system for maximum organic stabilization and minimum aeration requirement include the following: use of a bottom-mounted, self-aspirating aerator, an organic loading rate of around 15 Kg of biodegradable COD per $m^3$/day, with an aerator input of 0.18 kW/$m^3$ of reactor.

14. The autoheated air aerated sewage sludge digestion system would appear to be more economical than other aerobic or anaerobic digestion facilities at equal solids conversion efficiencies.

## SECTION 3

## RECOMMENDATIONS

1. Based upon the information developed by this study, a full scale thermophilic aerobic digestion facility can be designed for the organic stabilization and pasteurization of municipal sewage sludges and other waste organic slurries. The following approach is recommended for the design of a thermophilic aerobic digestion facility:

(a) the organic slurry that is to be digested should be characterized with respect to total chemical oxygen demand (TCOD) and solids concentration. TCOD value of the slurry would give an indication of the autoheating potential of the slurry since approximately 3.5 Kcal are released when 1 gram of TCOD is oxidized. A review of the history of the slurry, i.e., age, source, etc., would yield an approximate range of expected biodegradability.

(b) the reactor(s) should be sized according to the desired level of organic treatment efficiency and operating temperature, both of which are determined by the organic loading to the reactor. A loading rate of 12 to 15 kg TS/m$^3$ reactor-day was found to allow maximum temperature development (55 - 62°C) and maximum organic removal rate.

(c) the reactor should be shaped to provide efficient mixing, and should be covered and insulated. The insulation selected should provide maximum resistance to heat loss, specifically, the insulation should have a thermal conductivity of about 0.019 cal/sec · cm · °C, and be applied at a thickness to yield an $R > 25$.

(d) self-aspirating aerator(s) of the type studied here should be installed in the reactor(s) to provide approximately 0.15 - 0.20 kW/m$^3$ of reactor. This rate of aeration was found to provide efficient mixing and oxygen transfer. No other aerator should be substituted until it is shown to be capable of oxygen transfer efficiencies exceeding 20% in sewage sludge at 60°C.

(e) foam produced in the reactor(s) should be controlled with mechanical foam cutter(s) of the type studied here, and applied at the rate of 0.1 kW/m$^2$ surface (slurry) area. A backup foam cutter(s) should be available on an automatic activation basis for control of excess foam such as occurs during changes in loading conditions.

2. Information on the autoheated aerobic digestion process should be made available to engineering firms involved in the design of sludge treatment facilities where the ultimate disposal method would be land application. Effluent from the autoheated process was shown to be well stabilized and pasteurized, thereby minimizing the public health risks associated with land application of this type of sludge.

3. The potential of the autoheated aerobic digestion process for the solubilization of heavy metals, especially cadmium, and subsequent removal of the metals should be investigated further.

4. Additional studies concerning the dewatering characteristics of the thermophilic aerobic sludge should be conducted.

5. The potential for further improvement of the autoheated aerobic digestion process by the utilization of multistage reactors and heat exchangers should be investigated at the bench and full scale levels.

## SECTION 4

## BACKGROUND AND THEORY

The autoheating phenomenon commonly observed in composting of solids has not been applied to liquid organic slurries until recently. Due to the small amount of energy contained in waste organics such as sewage aludge and the large amount of water that would need to be heated, it was thought that autoheating might not be feasible with existing approaches to this topic. This study outlines a simple system and demonstrates how autoheating can be obtained with most waste organics using commercially available components. Although other studies have demonstrated the autoheating principle with sewage sludge, this study was the first large scale study to achieve autoheating to temperatures exceeding 60°C and to maintain this condition under continuous operation for more than a year. As will be shown, this was made possible primarily through the use of unique self-aspirating aerators.

### GENERAL SURVEY OF THE LITERATURE

### Effect of Temperature on the Rate of Oxidation

Although some conflicting information exists, it is generally agreed that increased temperatures increase the rate of substrate removal, and hence decrease the required reactor volume. The following simple model has been used to describe the substrate removal kinetics in a completely mixed system:

$$\frac{S_e}{S_o} = \frac{1}{1 + Kt} \qquad (1)$$

where $S_e$ = biodegradable organics [total volatile solids (TVS), gm/ℓ] in the effluent

$S_o$ = influent total biodegradable organics, TVS in gm/ℓ

$t$ = hydraulic retention time, day

$K$ = biodegradable removal rate coefficient, gm biodegradable TVS removed per gm of biodegradable TVS in the unit per day, or $\text{day}^{-1}$.

The substrate removal model is often adjusted to the temperature of operation by modifying K, the substrate removal rate coefficient, using the following equation derived from an Arrhenius relationship:

$$K_t = K_{20}\theta^{(T-20)} \tag{2}$$

where $K_{20}$ = removal rate at 20°C

$K_t$ = removal rate at other temperatures

$\theta$ = empirical factor relating the influence of temperature to the substrate removal rate over a specific temperature range.

When the value of $\theta$ is 1.017, this indicates that the removal rate doubles for every 10°C. Numerous observations have shown that biochemical reactions often double in rate for each 10°C rise in temperature for the values between 10°C and 35°C. This observation has resulted in a generalization called the "$Q_{10}$ rule."

If sewage sludge oxidation rates increased according to the $Q_{10}$ rule, the removal rate at 60°C would be 16 times as fast as it was at 20°C. It is doubtful that the common "rule of thumb" that biological processes double their microbial activity for every 10°C increase is valid up to a temperature of 60°C. Popel and Uhnmacht (11) reported that the organic substrate removal coefficient increased from 0.288 $\text{day}^{-1}$ to a value of 1.14 $\text{day}^{-1}$ as the temperature increased from the mesophilic range to 50°C. Gotaas (12) and Hawker et al.(13) noted that organic substrate removal rates double when the temperature increases from 10°C to 33°C, and increase fivefold over the 10°C rate at 53°C. Smith et al. (14) reported in a summary of sludge treatment around the country at different temperatures that the sewage sludge solids destruction rate coefficient increased from 0.15 $\text{day}^{-1}$ at 15°C to 0.37 $\text{day}^{-1}$ at 50°C. Matsch and Drnevich (15) reviewed the literature and estimated that the decay rate coefficient for several reported studies increased fourfold over a temperature range of 20°C to 60°C. It is likely that the influence of temperature is significantly altered by the species and variety of organisms present and the composition of the substrate.

## Oxygen Transfer and Autoheating

The development of a sludge treatment process capable of autoheating to thermophilic temperatures is dependent upon many factors, including substrate characteristics, reactor insulation, air temperatures, and so forth. However, it is generally agreed that the oxygen transfer capability of the aerator is the controlling parameter. Although the oxygen transfer coefficient increases with high temperatures due to lower liquid viscosities and density, high temperatures and high solids concentrations inhibit transfer efficiencies. The interaction of these parameters has not been defined, but the importance of oxygen transfer has been emphasized by several researchers (2,11,14,16,17, 18,19). Andrews and Kambhu (2) developed a computer model that combined an energy and mass balance for determining whether autoheated aerobic thermophilic sewage sludge treatment would be a feasible alternative treatment process. Figures 1 and 2 are reproduced from Andrews and Kambhu (2). The relative fate of the energy shown in Figure 1 illustrates the fate of the energy in these systems and indirectly emphasizes the importance of oxygen

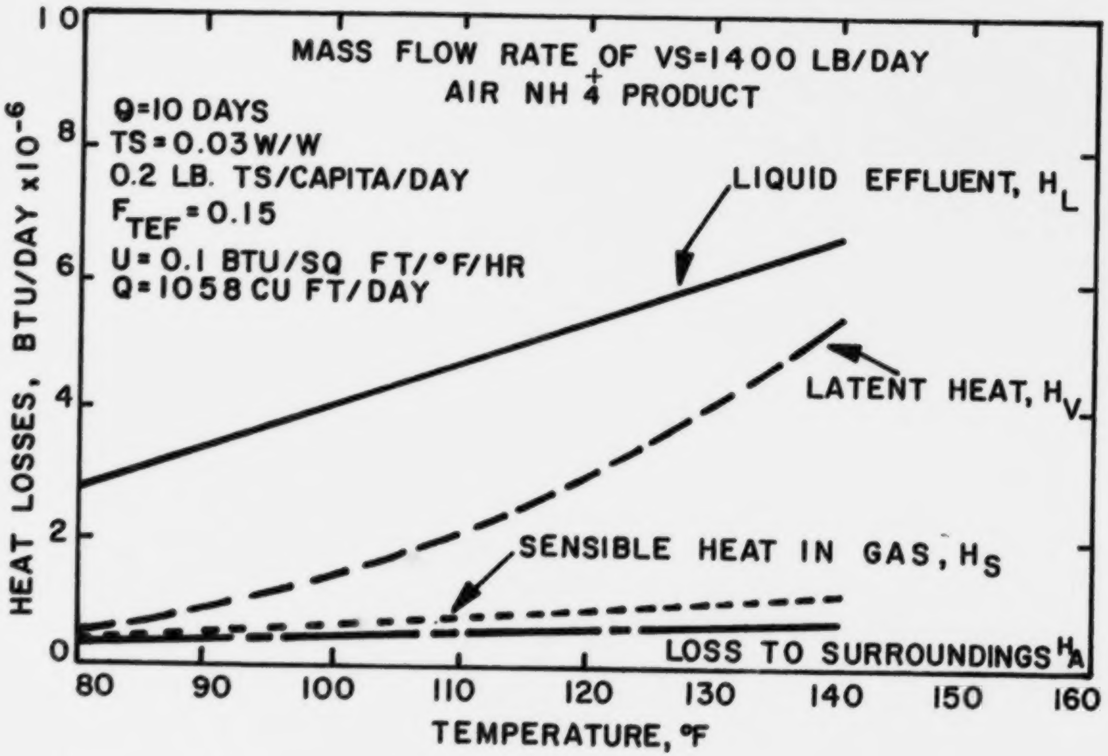

Figure 1. Computer model prediction of heat energy losses in aerobic digestion of sewage sludge (2).

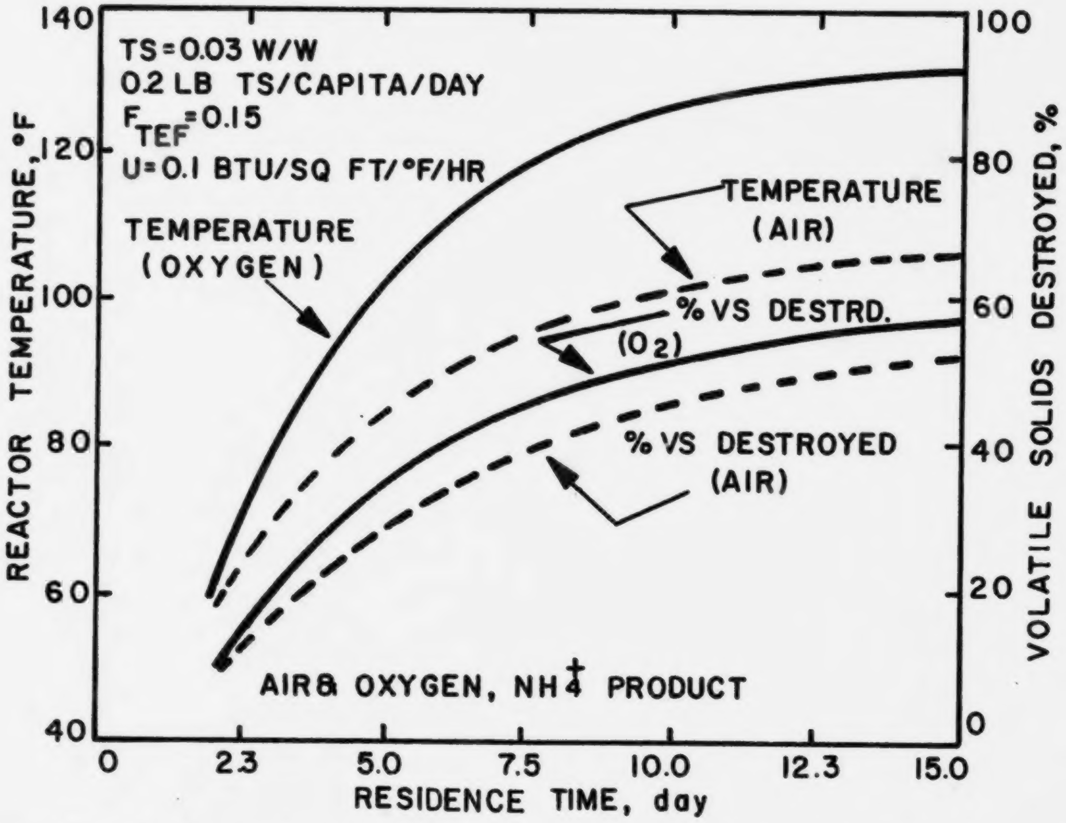

Figure 2. Comparison of the autoheating potential using air aeration and pure oxygen in sewage sludge treatment (2).

transfer. Of the total energy losses, the largest loss is in the heat contained in the effluent. At an operating temperature of 60°C (140°F) the energy lost through the heat of evaporation represents nearly half the heat that is lost and is nearly equivalent to that lost in the effluent waste. Note that these losses are predicted with the computer model for air aeration with a 15 percent oxygen transfer efficiency, a value far above the 3 to 5 percent values common for most aerators. The capability of achieving autoheating with air as compared to pure oxygen aeration is shown in Figure 2. The air aerated unit with a transfer efficiency of 15 percent did not achieve thermophilic temperatures, whereas the oxygen aerated unit was predicted to be capable of autoheating to greater than 50°C. The highly efficient air aerators evaluated in this study were not available for evaluation in this early review.

In a series of continuous feed laboratory studies, Wright (19) attempted to define a relationship capable of enabling prediction of operating digester temperatures in situations where the concentration of biodegradable organics and the volumetric loading rates were known. This laboratory study also predicted that thermophilic temperatures could be attained using air aeration with sewage sludge if transfer efficiencies greater than 15 percent could be achieved. Grant (20), in another study, concluded the following:

1. Utilizing air as the oxygen source, insufficient heat is generated biologically to compensate for energy losses to maintain thermophilic temperatures. A large energy input from mixing or some other source is needed to maintain thermophilic temperatures.
2. The necessary substrate-organisms-oxygen contact to produce rapid stabilization of organic material appears to be limited by the amount of oxygen that can be transferred from the air supply. Conventional equipment appears to be inadequate at thermophilic temperatures and high solids content to transfer the oxygen needed to maintain a high rate of biological activity.

Since conventional air aerators are capable of transfer efficiencies of 3 to 5 percent, it was reasonable to assume that autoheating of sewage sludge could only be achieved with oxygen supplemented air or pure oxygen. Smith et al. (14) and Matsch and Drnevich (15) demonstrated that this was feasible with pure oxygen. Recently, Matsch et al. (21) emphasized the above concerns.

A summary of experiences with autoheated aerobic digestion is shown in Table 1. Smith et al. (14) have summarized the details of most of these operations. Increases in temperature have ranged from zero up to 40°C with air and oxygen aeration of sewage sludge. Most of these studies have been short term with batch-type tests. Maximum autoheated temperature achieved with continuous full scale systems was 38°C.

The results of previous studies that have achieved limited autothermal conditions as shown, led a number of investigators to conclude that achieving thermophilic conditions via autothermal processes could only be accomplished under conditions that allowed efficient energy conservation, in particular, the use of pure oxygen to decrease the loss of energy through the latent heat of evaporation of water (15,21).

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TABLE 1. PREVIOUS LARGE SCALE EXPERIENCES WITH AUTOHEATED AEROBIC DIGESTION OF SEWAGE SLUDGE

| Type of Sludge | Oxygen Source | Size of Reactor ($m^3$) | Maximum Autoheated Temperature Achieved (°C) | Reference |
|---|---|---|---|---|
| Primary & Waste Activated | Pure Oxygen | 6,500 | 33 | 22 |
| Primary & Waste Activated | Air | Converted Full Scale Anaerobic Digesters | 38 | 23 |
| Primary | Air | 1.4 | 60 | 24 |
| Sewage Sludge | Air | 20 to 45 | 50 | 11 |
| Waste Activated Sludge | Oxygen | 6.4 | 44.5 | 14 |

Early European work on industrial wastes and animal wastes indicated that a simple self-aspirating aerator was capable of autoheating organic slurries to thermophilic temperatures (the empirical operating characteristics of a unique self-aspirating aerator using air, called the FUCH'S aerator was reported in the late 1960s (25)). Short term batch tests with various waste organics from industrial wastes and various animal raising operations reported autoheating to 50°C to 60°C in a number of cases (26). Subsequently, this aerator and several comparable designs were successfully marketed and implemented for industrial and agricultural waste treatment. However, the basic fundamentals regarding the process definition and limitations were empirical, and little information became available that would enable reliable predictive relationships to be developed for a waste organic such as sewage sludge. Hoffman and Crauer (27) introduced the process in the United States and Terwilliger and Crauer (28) reported additional results. Their information indicated that dairy, beef, and swine wastes could be treated by air aerated digestion, autoheated to thermophilic temperatures. DeLaval Company markets this as the "Licom" process. "Licom" is a contraction of the phrase "liquid composting." Little information on oxygen transfer efficiencies and mass and energy balances was developed to clarify the reasons why the autoheating phenomenon occurred and whether it would be useful in applications to sewage sludge treatment. The Licom process was the full scale process tested at the Binghamton-Johnson City sewage facility, along with the Midland-Frings aerator. A detailed description of the test facility is given in the Materials and Methods Section.

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## PREVIOUS AUTOHEATED AEROBIC DIGESTION STUDIES AT CORNELL UNIVERSITY

A research team at Cornell University has been attempting to define the requirements to achieve autoheated thermophilic aerobic digestion of common organic wastes since May, 1973. Only a brief summary of comments will be made in relation to those studies that have been completed (6,7) using animal wastes prior to start up of the sewage sludge study.

The basic objective of both the small and the full scale studies was to determine whether the heat released during microbial oxidation of organic residues would be sufficient to increase the temperatures of wet sludges (greater than 95% water) above 45°C without the addition of external heat energy. The laboratory studies began in May, 1973 and were terminated in December, 1974; full scale studies began in April, 1975. The substrate in most tests was composed of cow manure diluted to concentrations typical of many waste sludges. The concentration of feed substrate was varied from 2 to 12 percent volatile suspended solids. Cow manure has a maximum biodegradable fraction of about 40 percent of the total solids.

The small scale 30 liter reactor studies conducted by Wright (19) were successful in developing the relationship between decomposition kinetics and the potential for achieving autoheating to thermophilic temperatures. These results indicated that the conservation of heat generated during microbial oxidation would not be sufficient to achieve thermophilic temperatures using conventional air aerators. The large quantities of air required to satisfy the oxygen demand with conventional aerators resulted in the loss of most of the heat released during substrate oxidation. These data indicated that an oxygen transfer efficiency with air aeration greater than 15 percent would be required to achieve temperatures approaching 60°C with typical primary sludge.

Operation of the full scale two-stage (28.4 $m^3$ each) Licom system was preceded by extensive thermal and oxygen transfer studies. Characterization of the self-aspirating aerator using the conventional tap water analysis and comparison illustrated the lack of capability of predicting this kind of aerator performance. However, oxygen transfer measurements with the full scale system operating with cow manure indicated that efficiencies of oxygen transfer greater than 20 percent were being achieved in a 5 percent solids solution at 50°C.

This information emphasized the unique nature of this system and the need to carefully define its capabilities. Some of the data from the early studies with cow manure assist in illustrating the potential of the system. Figure 3 illustrates the relationship between the biodegradable organic loading rate (in kilograms of maximum biodegradable COD per cubic meter of reactor per day) to the oxygen transfer rate and efficiency. The optimum loading rate occurred at a value of about 9 kg biodegradable COD per $m^3$ of reactor per day with dairy cow manure. This loading rate is equivalent to a solids loading of about 0.4 lb VS per $ft^3$ per day for this particular organic slurry. The data in Figure 4 illustrate the relationship between the same organic loading rate, the reaction rate coefficient, and the removal rate. The organic loading rate of 9 kg BCOD/$m^3$-day correlates with the highest oxygen demand. Overall process efficiency, expressed as a percentage of the maximum BCOD that is destroyed, was about 60 percent at this optimum condition.

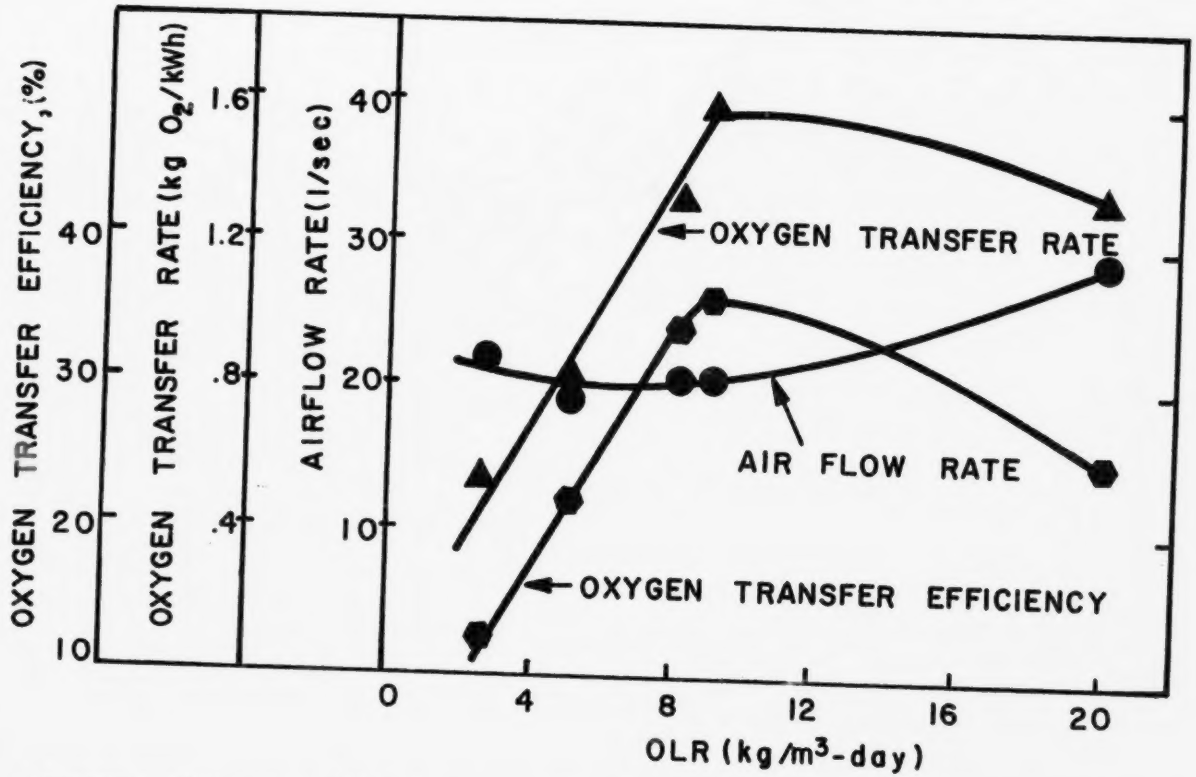


Figure 3. Aerator performance of the thermophilic digestions facility's self-aspirating aerator when treating cow manure diluted to 5% solids. OLR is the biodegradable COD loading rate (26).

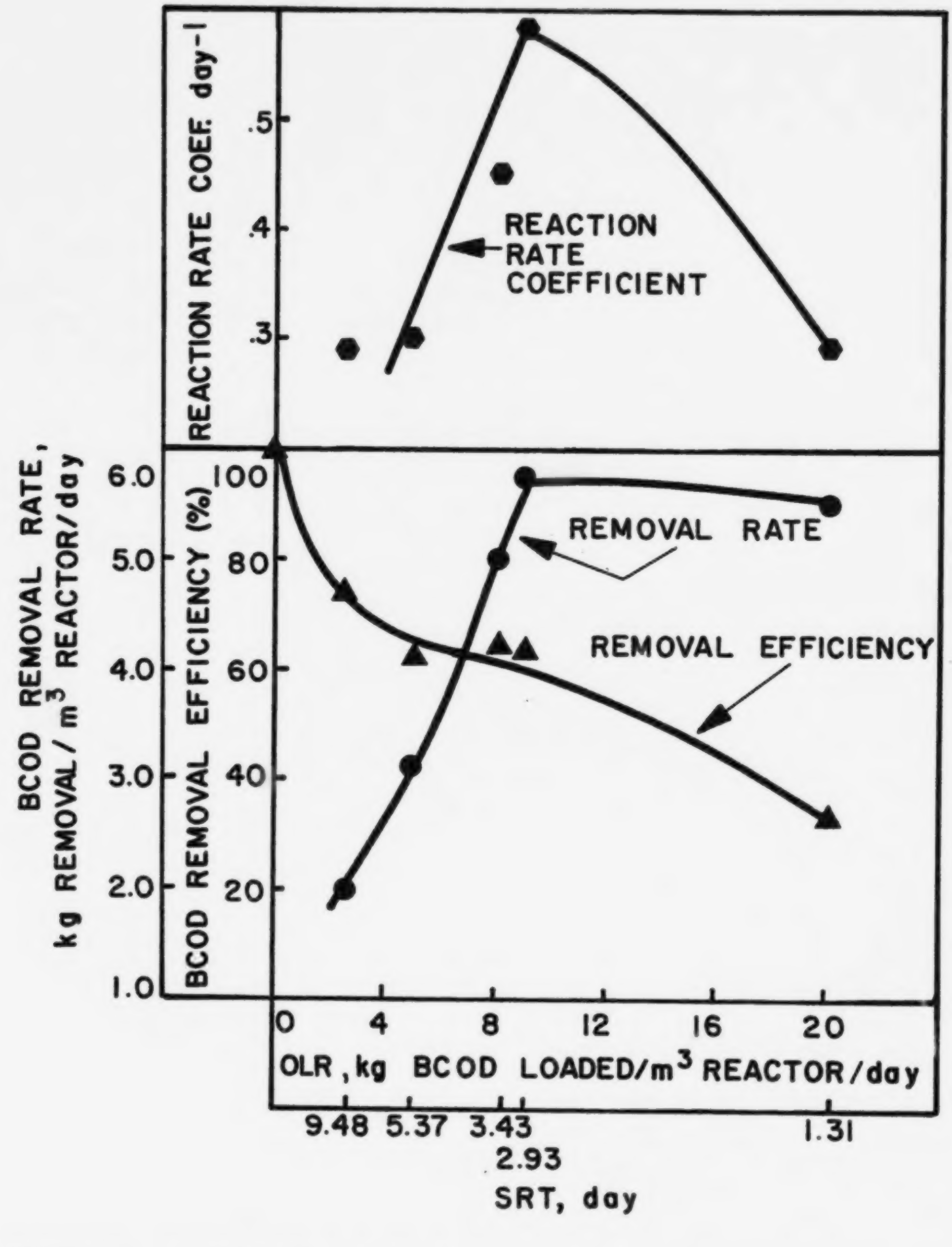


Figure 4. Relationship of loading rate, solids retention time, substrate removal rate and substrate removal efficiencies (26).

Due to the lack of understanding of the fundamentals affecting the DeLaval aerator, it was thought that the data shown in Figures 3 and 4 might not be directly translatable to sewage sludge. For example, maximum autoheating with a 5 percent cow manure slurry occurred at a biodegradable loading rate of about 9 kg BCOD/$m^3$-day, and 60 percent of the biodegradable solids were destroyed at this loading condition at a hydraulic retention time of 3 days. In order to achieve this loading rate with a 2 percent VS concentration sewage sludge, the hydraulic retention time would be about 2 days. This would require additional heat to maintain the reactor temperature at the increased flow rate. But the higher flow rate would decrease the solids destruction efficiency, thus decreasing the amount of energy released to the environment for autoheating.

## PATHOGENIC ORGANISMS IN SLUDGE

### Introduction and Overview

An individual human being sheds millions of microorganisms from his or her gastro-intestional tract each day into the sanitary sewer system. There, the organisms become an integral part of the stream of municipal wastewater that will eventually be treated and released to the environment. This municipal wastewater, typically speaking, has been shown to contain a wide variety, as well as a large number, of enteric microorganisms (29,30,31,32, 33,34,35). Many of these organisms are pathogenic for man and animal. Several studies have demonstrated that conventional wastewater treatment processes will remove many of these enteric microorganisms from the liquid fraction, especially if the final effluent is disinfected, e.g., with chlorine, bromine, ozone, etc., prior to discharge (36,37,38,39,40,41,42,43). However, many organisms become associated with the solids, i.e., sludge, yielding high concentrations of these organisms in both raw and secondary wastewater treatment plant sludges (44,45,46,47,48,49,50).

Traditional biological stabilization processes for sludge, such as aerobic or anaerobic digestion, have been shown to lower the number of pathogenic organisms in the sludge, but these processes do not achieve complete removal (disinfection or pasteurization) of the pathogens (4,50,51,52,53,54, 55). The literature also describes other methods of sludge treatment that act to reduce the levels of pathogens in sludge (56,57,58,59). These techniques include drying or dessication, heat treatment (pasteurization, etc.), irradiation, long-term storage, chemical treatment, and thermoradiation. With the direction of ultimate sludge disposal going toward land application, questions concerning the health risks of sewage sludge need to be investigated--especially questions concerning cost effective methods of sludge disinfection. In response to these concerns, research continues in the area of sludge disinfection with new emphasis on ionizing radiation, heat treatment (physically and biologically produced heat), and chemical treatment (58,59).

## Important Pathogenic Organisms

Three major groups of organisms are of primary concern in assessing the health risks of sludge. The groups contain organisms that are pathogenic for both man and animal. Some of the important members of each group are presented in Table 2. It has been well established that raw and digested municipal wastewater sludges contain a number of these organisms.

TABLE 2. PATHOGENIC ORGANISMS ASSOCIATED WITH WASTEWATER SLUDGES

| Bacteria | Virus | Parasites |
|---|---|---|
| 1. Enterobacteriaceae | 1. Enterovirus | 1. Protozoa |
| (a) Salmonella | 2. Adenovirus | (a) Entamoeba histolytica |
| (b) Shigella | 3. Agent Causing Infectious Hepatitis | 2. Helminths |
| (c) Klebsiella | | (a) Schistosoma mansoni |
| 2. Pseudomonads | | 3. Nematodes |
| 3. Clostridium | | (a) Ascaris sp. |
| 4. Mycobacterium | | |

### Bacteria--

The enteric bacteria, e.g., Salmonella, for the most part are known to be heat sensitive (thermolabile). Although the optimum growth temperature for many of the pathogenic (Enterobacteriaceae) organisms ranges from 35°C to 42°C, long exposure to temperatures above 45°C is lethal for these microbes. In fact, a study by Drnevich and Smith (60) showed that at a temperature of 45°C, Salmonella sp. (seeded into sewage sludge) were below detectable limits within 24 hours, while at 60°C, die-off of the microbes was quite rapid, with Salmonella sp. below detectable limits within several hours. The mechanism(s) of organism inactivation is protein denaturation at high temperatures in combination with the chemical harshness of the sludge environment, e.g., pH. An important factor in enteric pathogen survival in thermophilic digestion would be competition with thermophilic organisms for available substrate, nutrients.

### Viruses--

The study of virus occurrence in sludges and their inactivation in sludge treatment is currently a topic of much research. Ward and Ashley (52,57,61, 62,63) have published numerous papers of late concerning the mechanisms of virus inactivation. From their studies it may be generalized that at temperatures of 50°C or greater and high pH, i.e., pH $\geq 7$, virus inactivation within several hours will be assured. Another study performed in Europe on the fate of virus in the aerobic thermophilic digestion of animal manures also demonstrated the concept of heat and pH synergism for virus inactivation (64).

### Parasites--

One of the most environmentally resistive pathogenic organisms is the egg of the roundworm, Ascaris sp. Several studies have demonstrated the tenacity of these organisms (65,66,67). It is generally conceded that temperatures of 60°C for exposure times of at least 30 minutes are needed for inactivation of

*Ascaris*. Inactivation temperatures are somewhat lower for the environmentally more sensitive parasitic cysts, e.g., *Entamoeba*.

### Previous Studies: Fate of Pathogens During Aerobic Thermophilic Digestion

European researchers have long been concerned with the presence of pathogens in human and animal manures (64,68). Much research has been performed on the survival of virus and pathogenic bacteria in the liquid composting process (autothermal aerobic digestion) in Europe (64,68). The results have shown that the process eliminates virus and *Salmonella* sp. from the manures. Another study (60) on aerobic thermophilic digestion of sewage sludges found that the thermophilic digestion process successfully brought enteric pathogenic bacteria, i.e., *Salmonella*, below detectable limits with temperatures of 50°C and hydraulic retention times of one day. These studies on thermophilic digestion bear out earlier and on-going research on the fate of pathogens during composting (65,66,67,69). Composting was shown in several of these studies to effectively pasteurize the sludge with respect to bacteria, virus, and parasites with compost heap temperatures often above 60°C.

## THEORETICAL POTENTIAL OF AUTOHEATING

Aerobic biological oxidation of reduced organic compounds releases energy, part of which is used for microbial cell growth and maintenance. The conversion of organics to new cell material is not 100 percent efficient, and a portion of the total energy is released as heat into the environment. Since it is of great importance to quantify both the cell yield and the fate of metabolic products, considerable attention has been given to the fate of organics during microbial oxidation. Measures of cell growth or substrate oxidation have focused on the following: the number of adenosine triphosphate (ATP) units synthesized, the substrate carbon or hydrogen units converted, free energy change, and the theoretical oxygen equivalence (COD) of the converted organics. Servizi and Bogan (70) noted that the free energy was directly proportional to COD of the substrate and McCarty (71) showed this as the best approximation of the free energy changes involved in microbial growth. Since it is the easiest direct measure of energy change in a liquid slurry, this was the main parameter used in the study.

The relationship between energy loss to the environment as heat and that used for aerobic microbial cell growth is useful in illustrating the potential autothermal concept. The thermodynamics of biochemical conversions and the change in COD of oxidized organics provide the basis of the simplified modeling approach used in this study. The free energy values for substrate conversion can be calculated from the available values of free energy of formation if the stoichiometry of a reaction is known (71). This can be illustrated as follows with glucose oxidation:

$$C_6H_{12}O_6 + 6O_2 \rightarrow 6CO_2 + 6H_2O \tag{2a}$$

The free energy available during substrate utilization would be the sum of the free energy of formation of six moles of carbon dioxide plus six moles of water minus six moles of oxygen minus one mole of glucose, for a total of 688 Kcal per mole of glucose.

Equation (2a) indicates that one mole of glucose contains 192 grams of theoretical oxygen equivalence, or theoretical COD. Thus, the energy released per unit of glucose COD oxidized would be 3.59 Kcal per gram COD. A list of common organics and the relation of COD to free energy is given in Table 3.

TABLE 3. FREE ENERGY OF SUBSTRATE CONVERSIONS AS RELATED TO THE THEORETICAL COD FOR AEROBIC OXIDATIONS (71)

| Organic Compound | Free Energy of Conversion per gram COD |
|---|---|
| Glucose | 3.59 |
| Alanine | 3.29 |
| Pyruvate | 3.40 |
| Succinate | 3.22 |
| Formate | 3.77 |
| Acetate | 3.16 |

Cooney et al. (72) measured the heat produced by pure bacterial cultures utilizing various substrates and found that it equaled 3.44 ± 0.3 Kcal per gram of COD oxidized.

For many common organics, the quantity of energy that they contain is nearly a constant equal to 3.5 Kcal per gram COD or 8 to 12 Kcal per gram carbon. The maximum range of values for organic materials ranges from about 3 to 4 Kcal per gm COD, and this has been observed experimentally (7,19).

The relationship between energy and organic conversion efficiencies are schematically illustrated in Figure 5 using glucose as an example. The following reaction illustrates the chemical conversions during initial biological oxidation of a simple sugar:

$$C_6H_{12}O_6 + O_2 + NH_3 \rightarrow CO_2 + 4H_2O + C_5H_7O_2N \qquad (3)$$

Rapid bacterial growth will result in the synthesis of 113 grams of bacteria ($C_5H_7O_2N$) with the consumption of one mole of glucose and one mole of oxygen. Of the total free energy (688 Kcal per mole of glucose oxidized), about 560 Kcal remains as chemical energy in the new bacterial mass; the difference, or 128 Kcal, would be released as heat to the environment. Although this example is indicative of an exceptionally high energy conversion efficiency to new cells, endogenous respiration would rapidly release much of the free energy contained in the new cells. The efficiency of energy conversion under aerobic conditions from different substrates varies from 12 to 82 percent, with many being less than 30 percent efficient (71). Much of the energy of conversion of organics results in addition of heat to the environment; and when enough organic matter is present, this may result in temperature increases.

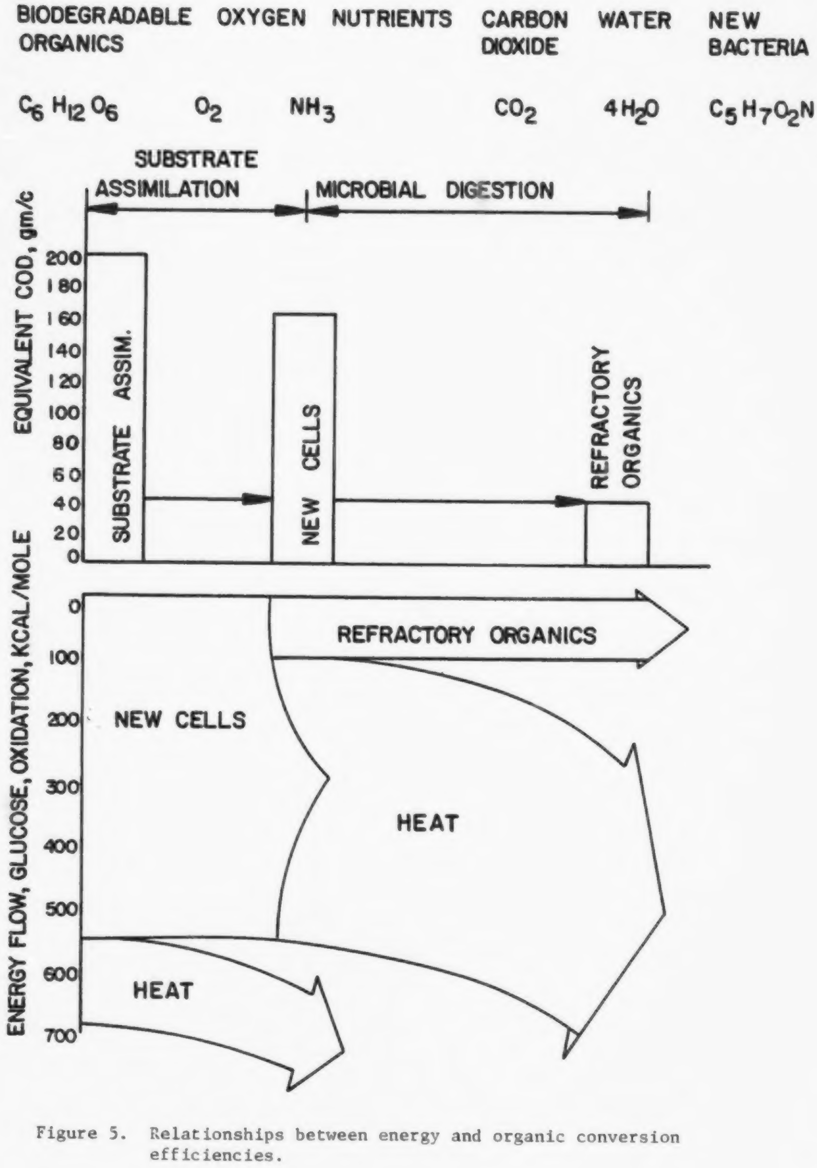

BIODEGRADABLE ORGANICS
OXYGEN
NUTRIENTS
CARBON DIOXIDE
WATER
NEW BACTERIA
$C_6H_{12}O_6$
$O_2$
$NH_3$
$CO_2$
$4H_2O$
$C_5H_7O_2N$
SUBSTRATE ASSIMILATION
MICROBIAL DIGESTION
EQUIVALENT COD, gm/c
200
180
160
140
120
100
80
60
40
20
0
SUBSTRATE ASSIM.
NEW CELLS
REFRACTORY ORGANICS
ENERGY FLOW, GLUCOSE, OXIDATION, KCAL/MOLE
0
100
200
300
400
500
600
700
REFRACTORY ORGANICS
NEW CELLS
HEAT
HEAT

Figure 5. Relationships between energy and organic conversion efficiencies.

Thus, because of the relative constant relationship between heat released during oxidation of organics as shown in Table 3, the total biological heat that can be generated can be estimated using the following relationship:

$$\Delta F = 3.5\ \Delta COD \qquad (4)$$

where $\Delta F$ = the total heat released in Kcal/ℓ

$\Delta COD$ = the measured change in the COD in gm/ℓ

These units are convenient in estimating liquid temperature change since the heat released, expressed as Kcal per liter, is equal to the temperature change (i.e., one Kcal is about one°C/liter, by definition). Although this introduces some error, since the thermodynamics vary depending upon the organic composition, the variation is probably less than 10 percent.

## MODELING THE AUTOHEATING POTENTIAL FOR SEWAGE SLUDGE

An overview of the total energy that can be made available in sewage sludge indicates the lower concentrations that were considered in this testing. Since some activated sludges can only be thickened to about 1 percent, this was considered as the practical lower limit. Since it is known that about 3.5 Kcal of heat energy is released during the oxidation of one gram COD, the total available energy can be estimated and compared to various system losses. In a 1 percent (TVS) solids sludge that is 50 percent biodegradable, the total amount of heat energy that will be released can be estimated to be 24.5 Kcal per liter. The sensible heat required to raise the sludge from an influent temperature of 15°C to a thermophilic operating temperature (minimum of 43°C) is about equal to the exothermic heat of substrate oxidation. Thus, if this energy could be conserved, most sludges would contain enough energy to autoheat to the thermophilic temperatures. This would require an efficient heat exchanger with dilute sludges.

In order to determine the limitations of actually achieving the autoheated temperatures, a total systems mass and energy balance would be necessary. Although development of a systems model was not a major objective, a simplified model was developed. An energy balance accounting for the flows illustrated in Figure 6 was used as modified from earlier studies (7,19).

Each term affecting the operation of the system was examined in a steady-state heat balance of the aerobic thermophilic reactor. In addition, the heat balance could be reduced to a temperature balance, which would enable prediction of the steady-state reactor temperature.

The heat balance was developed using basic thermochemical and kinetic relationships and assumed the reactor was aerated continuously, completely mixed without recycle, continuously fed, and operating at steady state conditions. In general, the steady state heat balance is expressed by Equation (5).

$$\begin{matrix}\text{HEAT}\\ \text{INPUT}\end{matrix} = \begin{matrix}\text{HEAT}\\ \text{OUTPUT}\end{matrix} - \begin{matrix}\text{THE RATE OF}\\ \text{BIOLOGICAL HEAT PRODUCTION}\end{matrix} \qquad (5)$$

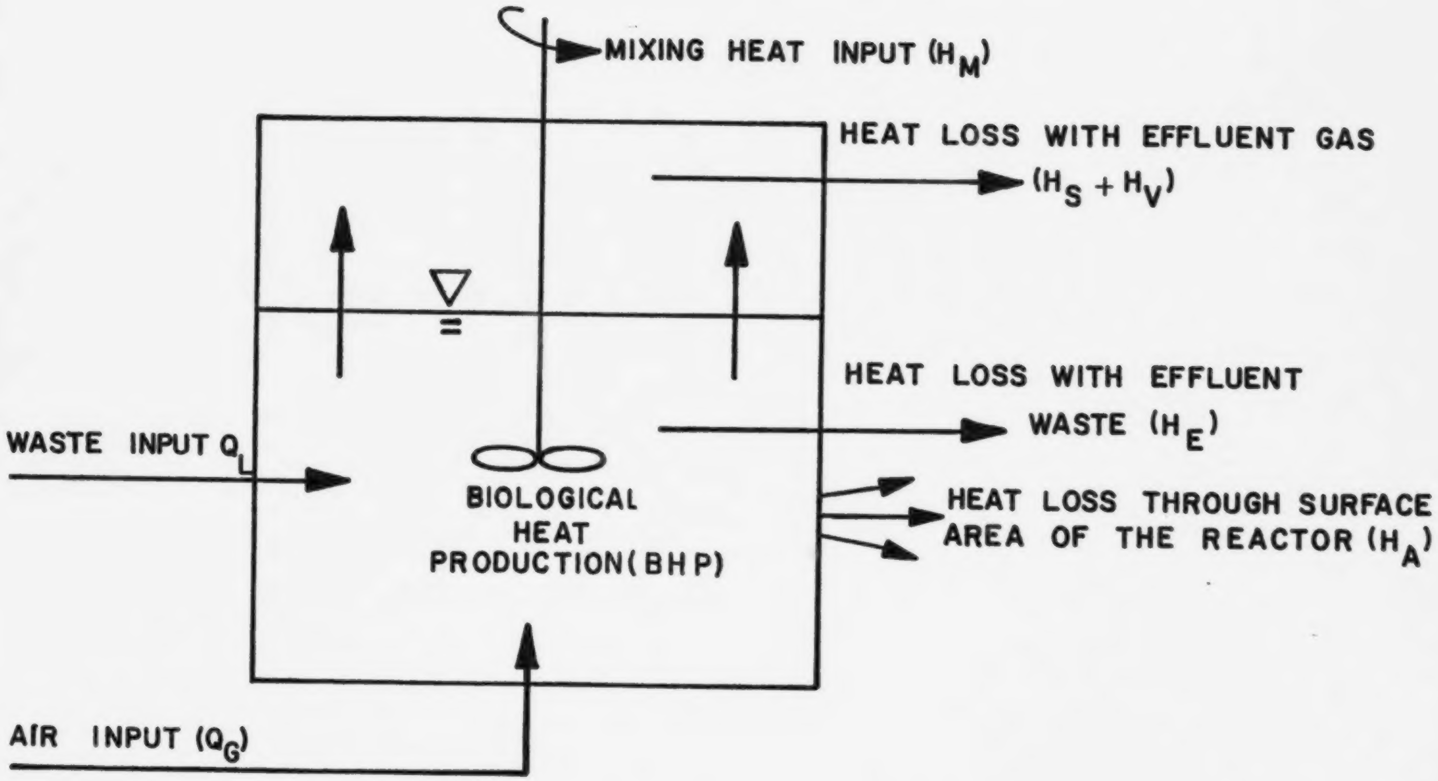

MIXING HEAT INPUT ($H_M$)
HEAT LOSS WITH EFFLUENT GAS ($H_S + H_V$)
HEAT LOSS WITH EFFLUENT WASTE ($H_E$)
WASTE INPUT $Q_L$
BIOLOGICAL HEAT PRODUCTION(BHP)
HEAT LOSS THROUGH SURFACE AREA OF THE REACTOR ($H_A$)
AIR INPUT ($Q_G$)

Figure 6. Illustration of parameters considered in the energy balance model of a completely mixed biological reactor (26).

The major heat inputs are the biological heat production (BHP) and the heat input due to the mechanical mixing of the self-aspirating aerator ($H_M$). The heat outputs are heat loss in the effluent waste ($H_E$), heat loss as latent heat in water vapor ($H_V$), heat loss as sensible heat in the effluent gas ($H_S$), and the heat loss through the surface area of the reactor ($H_A$). A schematic of the reactor illustrating the terms considered in the heat balance is presented in Figure 6. The overall heat balance may now be more specifically defined by Equation (6).

$$BHP + H_M = H_E + H_V + H_S + H_A \qquad (6)$$

Many of the terms presented in this equation are difficult to measure or calculate. For this reason, preliminary tests were performed on the reactor to simultaneously determine $H_M$, $H_V$, $H_S$, and $H_A$. These preliminary tests consisted of monitoring the decrease in reactor temperature while the aerator was operating in hot tap water. A system specific energy balance (Equation (7)) was developed from these empirical data, and a heat loss coefficient ($H_L$) was defined which would allow the calculation of the heat loss as a result of these terms.

$$H_L = H_V + H_S + H_A - H_M \qquad (7)$$

$$H_L = a(T_R - T_A)^b D_L C_{PL} \qquad (7a)$$

where $T_R$ = reactor temperature, °C
$T_A$ = ambient air temperature, °C
$D_L$ = liquid density, gm/ℓ
$C_{PL}$ = specific heat capacity of the liquid, cal/gm-°C
a = intercept coefficient
b = slope coefficient.

The temperature of the effluent is theoretically the same as that of the reactor for a continuously flowing system. The heat loss due to the effluent waste ($H_E$) is therefore equal to the quantity of heat required to raise the temperature of the influent waste to the reactor operating temperature. This heat loss can be calculated with Equation (8).

$$H_E = \frac{T_R - T_o}{HRT} D_L C_{PL} \qquad (8)$$

where $T_o$ = feed temperature, °C
HRT = hydraulic retention time, day.

A mass balance (Equation (9)) was utilized to establish the steady-state kinetics of the system, which assumed the rate of removal of the biodegradable portion of the total chemical oxygen demand (BCOD) was first order with respect to the BCOD.

$$Q_L(BCOD_o) - Q_L(BCOD_e) = (d\ BCOD_e/dt)V \qquad (9)$$

where $Q_L$ = liquid flow rate, volume/time

$BCOD_o$ = biodegradable COD in the influent

$BCOD_e$ = biodegradable COD in the effluent

$V$ = system volume.

Therefore, by combining terms and simplifying by integration, the BCOD in the effluent can be calculated with Equation (10).

$$BCOD_e = BCOD_o/1 + K_T HRT \qquad (10)$$

where $K_T$ = reaction rate coefficient, $day^{-1}$

HRT = hydraulic retention time, day.

Since the biological heat production is proportional to the rate of BCOD destruction, it can be calculated using Equation (11).

$$BHP = K_T\ BCOD_e\ R \qquad (11)$$

where R = the quantity of heat released to the surroundings by the bacteria is Kcal per gram COD removal (this value has been shown to vary between 3 and 4 Kcal/gram COD removal depending on the growth rate of the microbes in the system).

The overall steady-state heat balance can now be expressed by Equation (12).

$$BHP = H_L + H_E \qquad (12)$$

The overall steady-state temperature balance can be determined by expanding the terms in Equation (12) as shown below.

$$K_T\ BCOD_e\ R = a(T_R - T_A)^b\ D_L C_{PL} + \frac{D_L C_{PL}(T_R - T_o)}{HRT} \qquad (13)$$

In summary, the quantity of heat released to the liquid sewage slurries during aerobic biological oxidation is theoretically sufficient to increase the temperature to thermophilic levels if it can be conserved within the reactor. However, conventional air aeration systems strip most of this heat unless efforts are made to conserve it. A recent study (7) indicates that self-aspirating aerators that utilize air as an oxygen source are efficient enough to conserve the necessary amount of energy to support autoheating in insulated reactors.

Thus, the general requirements for autoheated thermophilic digestion with air aeration appear to be:

-sufficient biodegradable organics to provide heat of oxidation equal to 25 to 35 Kcal per liter, depending upon the incoming sludge temperature;

-insulated reactor to conserve reactor heat; and

-oxygen transfer efficiency exceeding 10 percent.

# SECTION 5

# INVESTIGATIVE FACILITIES, BINGHAMTON-JOHNSON CITY SEWAGE TREATMENT PLANT OPERATION, AND SYSTEM DESIGN AND CONSTRUCTION

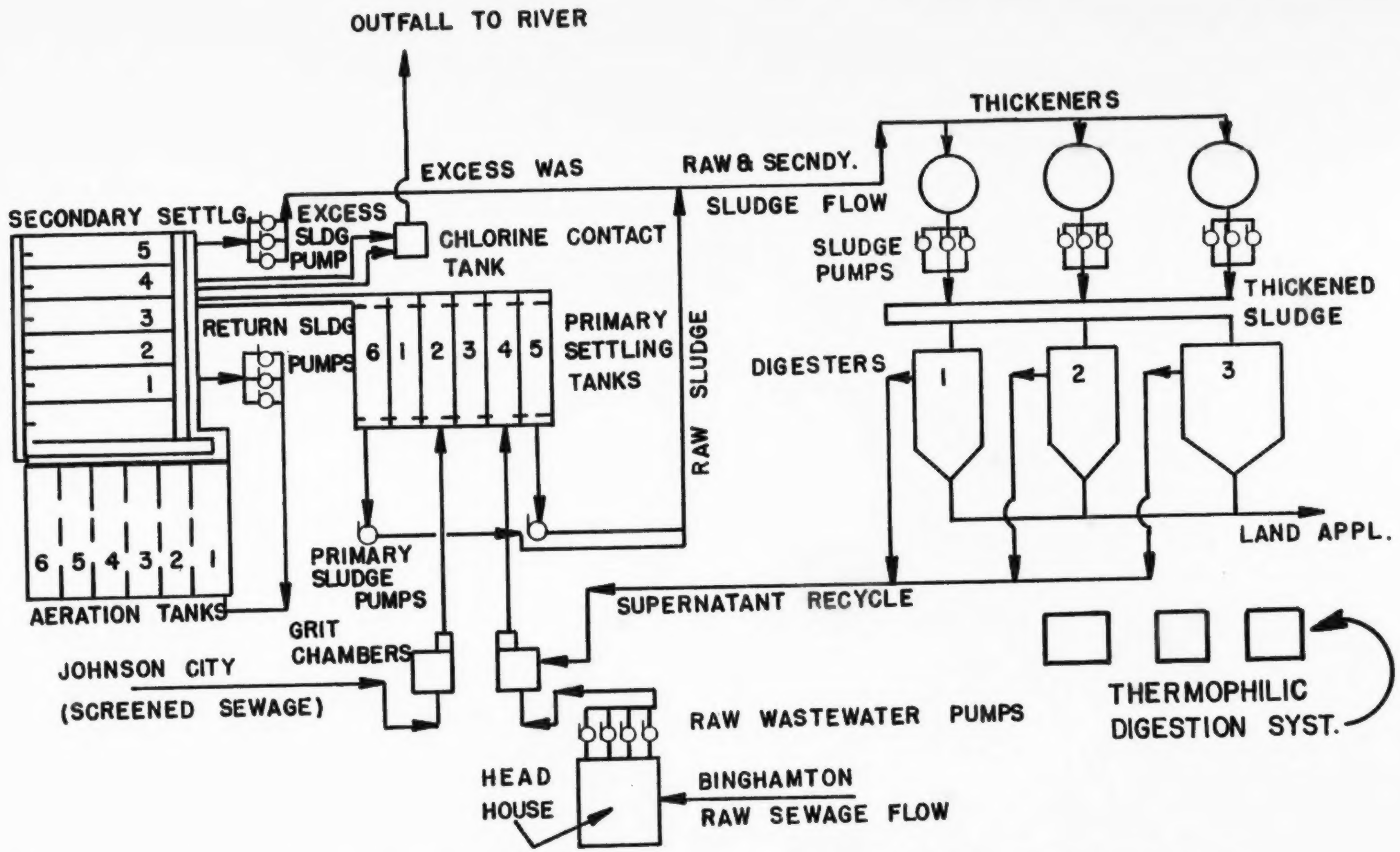
## INVESTIGATIVE FACILITIES

The full scale demonstration study was conducted at the Binghamton-Johnson City Sewage Treatment Plant (STP) located in Binghamton, New York (Broome County). This facility was constructed in the 1950s and has been upgraded several times to handle the increase in sewer connections as well as to meet revised discharge standards. The actual operation of this plant will be discussed in a subsequent section. A diagram showing the relative placement of the sewage treatment plant's major components and the location of the Cornell University digestion system is presented in Figure 7.

The Binghamton-Johnson City STP is under authority of the Binghamton-Johnson City Sewage Treatment Board with the operational responsibilities delegated to the plant superintendent, Mr. Eugene F. Martiny, P.E., and the senior operator, Mr. Jack P. Bradley. Mr. Steve Cerny, the plant chemist, conducted physical and chemical analyses for monitoring of STP performance. These analyses were performed in the STP wet chemical laboratory located in the Head House. The day-to-day operational duties were performed by the three working shifts of operators and mechanics.

During the first nine months of the study, the waste management laboratory at the Cornell Animal Science Teaching and Research Center (ASTARC) served as the support facility for this investigation. Analyses of the sludge samples from the full scale system, as well as bench scale studies, were executed in this facility. The Cornell research team associated with this project was also located at the waste management laboratory.

For the duration, i.e., 15 months, of the study, the Agricultural Waste Management Laboratory (AWML) at Cornell University served as the support facility for this investigation. The laboratory contains a wet chemical analyses area, pilot plant process equipment, and animal housing facilities. A floor plan of this facility is illustrated in Figure 8. The Cornell personnel associated with the project were located at the AWML. Analyses of all sludge samples in addition to various supplemental studies were conducted in this facility.


OUTFALL TO RIVER
EXCESS WAS
RAW & SECNDY.
SLUDGE FLOW
THICKENERS
SECONDARY SETTLG
5
4
3
2
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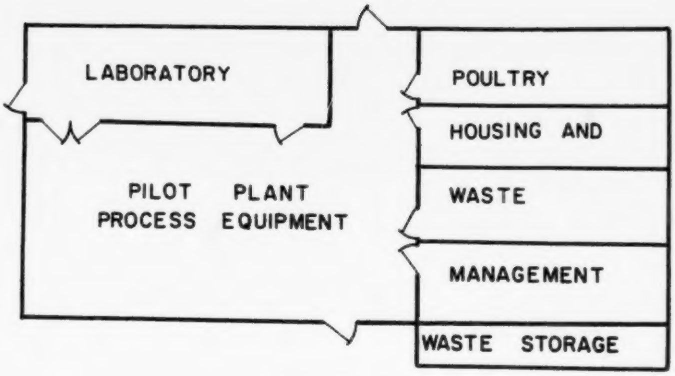
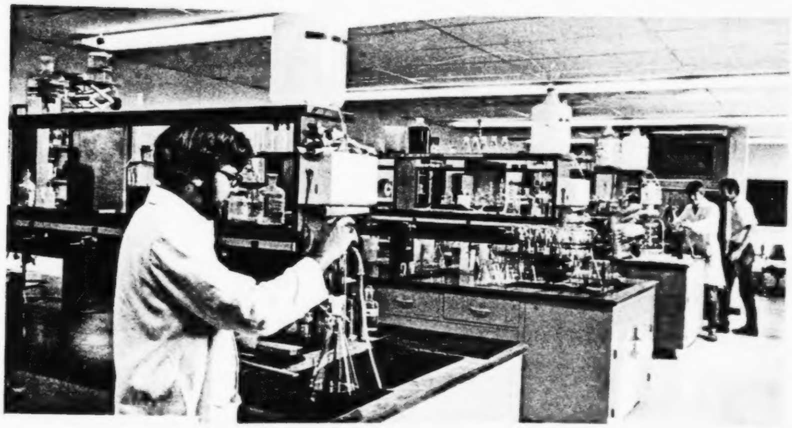
EXCESS
SLDG
PUMP
CHLORINE CONTACT
TANK
SLUDGE
PUMPS
THICKENED
SLUDGE
RETURN SLDG
PUMPS
6
1
2
3
4
5
PRIMARY
SETTLING
TANKS
RAW SLUDGE
DIGESTERS
1
2
3
6
5
4
3
2
1
AERATION TANKS
PRIMARY
SLUDGE
PUMPS
LAND APPL.
SUPERNATANT RECYCLE
GRIT
CHAMBERS
JOHNSON CITY
(SCREENED SEWAGE)
RAW WASTEWATER PUMPS
THERMOPHILIC
DIGESTION SYST.
HEAD
HOUSE
BINGHAMTON
RAW SEWAGE FLOW

Figure 7. An aerial view diagram illustrating the Binghamton-Johnson City STP components and relative location of Cornell University's digestion facility.


LABORATORY
PILOT PLANT PROCESS EQUIPMENT
POULTRY
HOUSING AND
WASTE
MANAGEMENT
WASTE STORAGE

FLOOR PLAN



ANALYTICAL LABORATORY

Figure 8. Floor plan and analytical laboratory in the Cornell Agricultural Waste Management Laboratory.

## BINGHAMTON-JOHNSON CITY SEWAGE TREATMENT PLANT OPERATION

### Introduction

The influence of sludge composition, i.e., concentration, volatility, etc., on the aerobic autothermal process, as will be shown in a subsequent section, was significant; therefore, it was important that the sludge generation at the sewage treatment plant be investigated. A description of the existing facilities, the composition of the influent wastewater, performance of the existing units, and the generation-characterization of the sludge is contained in this section. Much of this information was gleaned from the treatment plant records; however, a large amount of information was taken from the recent upgrade study on the sewage treatment plant as performed by Calicerinos and Spina, Consulting Engineers (Liverpool, N.Y.).

### Existing Facilities

The Binghamton-Johnson City STP was originally built in the 1950s; however, several expansions (upgrades), the last one in 1973, have brought the plant through secondary treatment. The major design parameters are as follows (74):

Flow ($m^3$/day)

Average = 69,076

Minimum = 22,710

Maximum = 138,153

Loadings (kg/day)

$BOD_5$ = 15,203

TSS = 16,585

A brief description of the existing treatment facilities is presented here and depicted graphically in Figure 7. Design specifications of the major process units are given in Table 4.

Primary clarification was designed based on an average daily influent flow of 69,076 $m^3$/day. Surface overflow rate (SOR) was to be 40.7 $m^3/m^2$-day at the average daily flow. The design depth was 3.35 m, thereby yielding a 2-hr hydraulic retention time.

The activated sludge process was used for secondary treatment. The treatment system was designed for operation in the complete-mix mode or the contact-stabilization mode giving the plant a large degree of flexibility. Under complete-mix operation, a hydraulic retention time of 3 hr was obtained at the design flow. In the contact-stabilization mode, contact and reaeration times provided were 0.75 hr to 1.8 hr and 1.4 hr to 11.4 hr, respectively.

Secondary clarification was designed for a SOR of 32.6 $m^3/m^2$-day at the design influent flow of 69,076 $m^3$/day. At a design depth of 3.35 m, the resulting hydraulic retention time was 2.7 hr.

Sludge treatment consisted of anaerobic digestion preceded by gravity thickening of the sludge. Gravity thickeners were designed for a maximum loading of 39.1 kg/$m^2$-day with a SOR of 24.4 $m^3/m^2$-day. The thickened sludge

## TABLE 4. DESIGN SPECIFICATIONS OF MAJOR PROCESSES UNITS

**Primary Clarifier**

Six basins, 14 Bays (4.57 m x 26.5 m x 3.66 m, each)

Surface area (total) = 1,695 $m^2$

Surface overflow rate = 40.7 $m^3/m^2$-day

Hydraulic retention time = 2.0 hr

Sludge collection: chain and flight

**Aeration Basins**

Six covered basins (9.14 m x 30.5 m x 5.49 m, each)

Total volume = 9,174 $m^3$

Hydraulic retention time = 1.6 - 2.5 hr

**Air System**

Four (Roots) centrifugal blowers (448 kW; 20,727 $m^3$/hr, each)

Design air requirement = 124.7 $m^3$ air/kg $BOD_5$

Swing arm aerators with fine bubble sock diffusers in Bays 1-4 & 6

Coarse bubble diffusers in Bay 5

**Secondary Clarifier**

Five basins (10.1 m x 42.1 m x 3.51 m, each)

Surface area (total) = 2,115 $m^2$

Surface overflow rate = 32.1 $m^3/m^2$-day

Hydraulic retention time = 1.3 - 2.2 hr

Sludge collection: chain and flight

**Thickeners**

Three: 1-10.7 m diameter; 2-15.2 m diameter; all with 3.1 m sidewall

Total surface area = 271 $m^2$

Solids loading = 39.1 kg/$m^2$-day

**Anaerobic Digesters**

Three: 1-21.3 m diameter; 2-15.2 m diameter; all with 9.6 m sidewall

Total volume = 6,524 $m^3$

Type: high rate; gas mix; external gas heat was to be fed to the anaerobic digesters, designed for high-rate stabilization, at an average loading of 1.92 kg volatile suspended solids/m$^3$-day. Since the gravity thickeners were designed to provide an average sludge concentration of 6% total solids, a hydraulic retention of approximately 21 days would be obtained in the anaerobic digesters. The digested sludge was used for land application purposes or recycled through the treatment plant which will be shown to be a limiting factor in the plant's performance.

## Wastewater Characterization

Wastewater flow to the Binghamton-Johnson City STP was highly variable, which was a function of the combined sewer system serving the community. Flow to the plant increased dramatically with precipitation, a direct result of storm drain connections and infiltration. Average wastewater flow for the period January, 1976 to January, 1978 was 68,727 m$^3$/day, with a range of 40,542 m$^3$/day to 159,197 m$^3$/day. Exfiltration also played a role in flow variation to the treatment plant during dry periods. Figure 9 shows the daily flow variation (averaged weekly) during the period May, 1977 to August, 1978.

Daily flow variations were investigated by Calcerinos & Spina. The results indicated that the total peak daily dry weather flow was 1.35 times the total average daily flow. During this same time period (dry weather), the minimum daily flow was 0.61 times the average daily flow.

Table 5 presents a summary of the wastewater flows for three categories, exclusive of infiltration and inflow.

TABLE 5. WASTEWATER FLOWS FOR THE BINGHAMTON-JOHNSON CITY STP (74)

| Wastewater Flows | Flow (m$^3$/day) |
|---|---|
| Residential | 23,846 |
| Commercial & Public | 9,614 |
| Industrial | 7,911 |
| Total Flow (Avg.), m$^3$/day | 41,370 |

A comparison of the table presented here with the total flows to the treatment plant readily shows the influence of inflow (storm water) and infiltration on flow variation.

Loadings to the treatment plant for BOD$_5$ and total suspended solids (TSS) (averaged over a two year period from January, 1976 to January, 1978) were 7,188 kg/day and 6,984 kg/day, respectively. A further breakdown of the loadings with regard to the source and composition of the source discharge was computed by Calcerinos & Spina (dry weather averages) and is presented in Table 6.

The discrepancies between the actual BOD$_5$ and TSS loadings to the plant over the past two years and the loadings computed by Calcerinos & Spina (see Table 6) were due to losses of BOD$_5$ and TSS through storm water overflows (or exfiltration) that discharged directly to the river. As with the flow to the

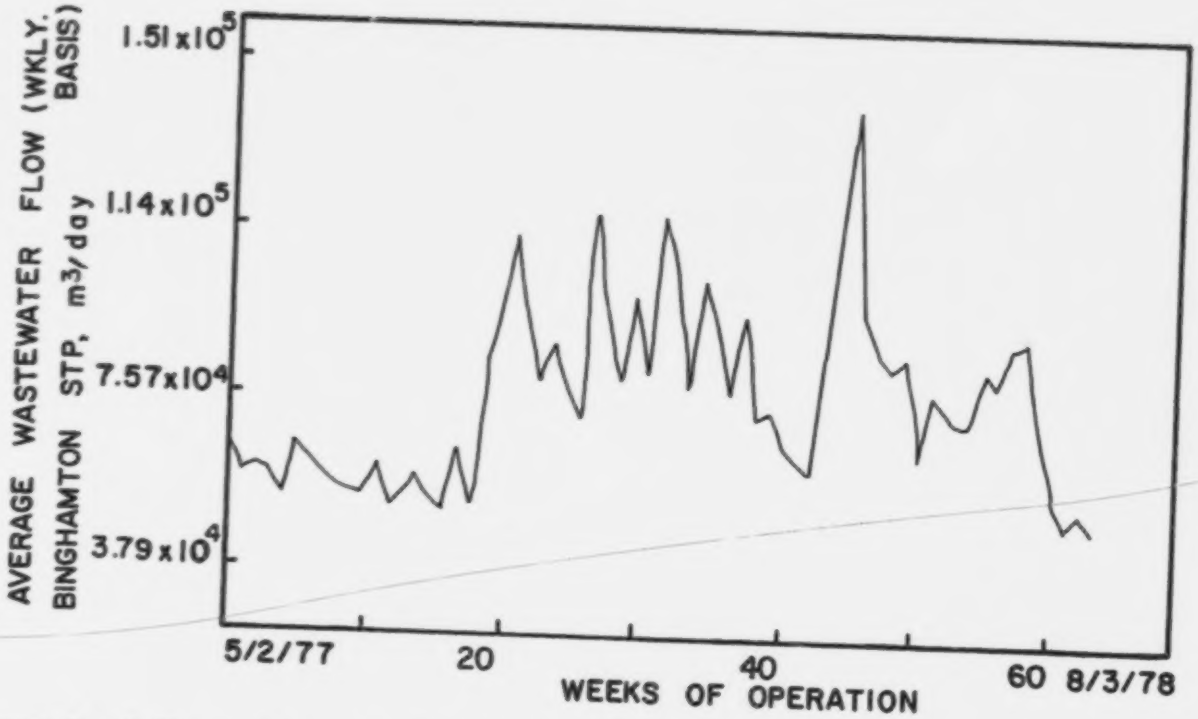


Figure 9. Influent wastewater flow variations during the period of full scale study.

TABLE 6. SOURCE AND COMPOSITION OF WASTEWATER DISCHARGED TO THE BINGHAMTON-JOHNSON CITY STP (74)

| Parameter (kg/day) | Residential, Commercial and Public | Industrial | Total |
|---|---|---|---|
| $BOD_5$ | 6,643 | 5,362 | 12,005 |
| TSS | 6,912 | 2.059 | 8,971 |
| TKN | 2,394 | 402 | 2,796 |
| $NH_3$ | 1,378 | 129 | 1,507 |

treatment plant, loadings were highly variable. $BOD_5$ loadings ranged from 1,908 kg/day to 22,178 kg/day, whereas TSS loadings were more highly variable ranging from 1,156 kg/day to 58,785 kg/day. (The average $BOD_5$ and TSS loadings are given above.)

## Performance Summary

Loadings to the treatment plant were variable, as was shown earlier. Figures 10 through 12 demonstrate the variability of $BOD_5$, TSS, and COD concentrations in the plant influent. As would be expected with a highly variable influent, similar fluctuations were observed in the effluent from the primary and secondary systems as seen in Figures 10 through 12. These fluctuations observed for the primary and secondary system performance were a function of the influent flow and concentration variability and, therefore, volumetric loading rate variability (Figure 13) as well as sludge recycle from the anaerobic digester(s) to the influent manhole (headworks). In fact, it was estimated that the sludge recycle flows (average for the two years of study) were approximately 170 $m^3$/day at a total solids concentration from 105 gm/$m^3$ to 180 gm/$m^3$ (74). It is pertinent to note that no records were kept of sludge recycle flow and concentration and that the numbers given here were drawn from a mass balance analysis of the entire treatment plant; therefore, all of the plant performance data are in terms of influent concentrations and flows excluding sludge recycle.

A summary of the major process units performance, i.e., averages with associated ranges, during the period of this study, is given in Table 7. For illustrative purposes, Figure 14 shows a schematic diagram of the Binghamton-Johnson City STP with each unit's average performance data.

## Sludge Generation and Characterization

Sludge was originated from the primary clarifier underflow and secondary clarifier underflow, then pumped to the sludge thickeners together. Sludge from the primary tanks averaged approximately 1% TSS, with a flow of 3,028 $m^3$/day (see Figure 15), while the secondary clarifier underflow averaged approximately 0.5% TSS, with a flow (excess) ranging from 0 $m^3$/day to 1,222 $m^3$/day (Figure 15). The mixed liquor suspended solids (MLSS) concentration was the controlling parameter for determination of the amount of sludge excessed from the secondary clarifiers. MLSS concentration was kept at a low level due to the limited (clarifying) ability of the secondary clarifiers. Figure 15 illustrates the raw, return, and excess sludge flows during the

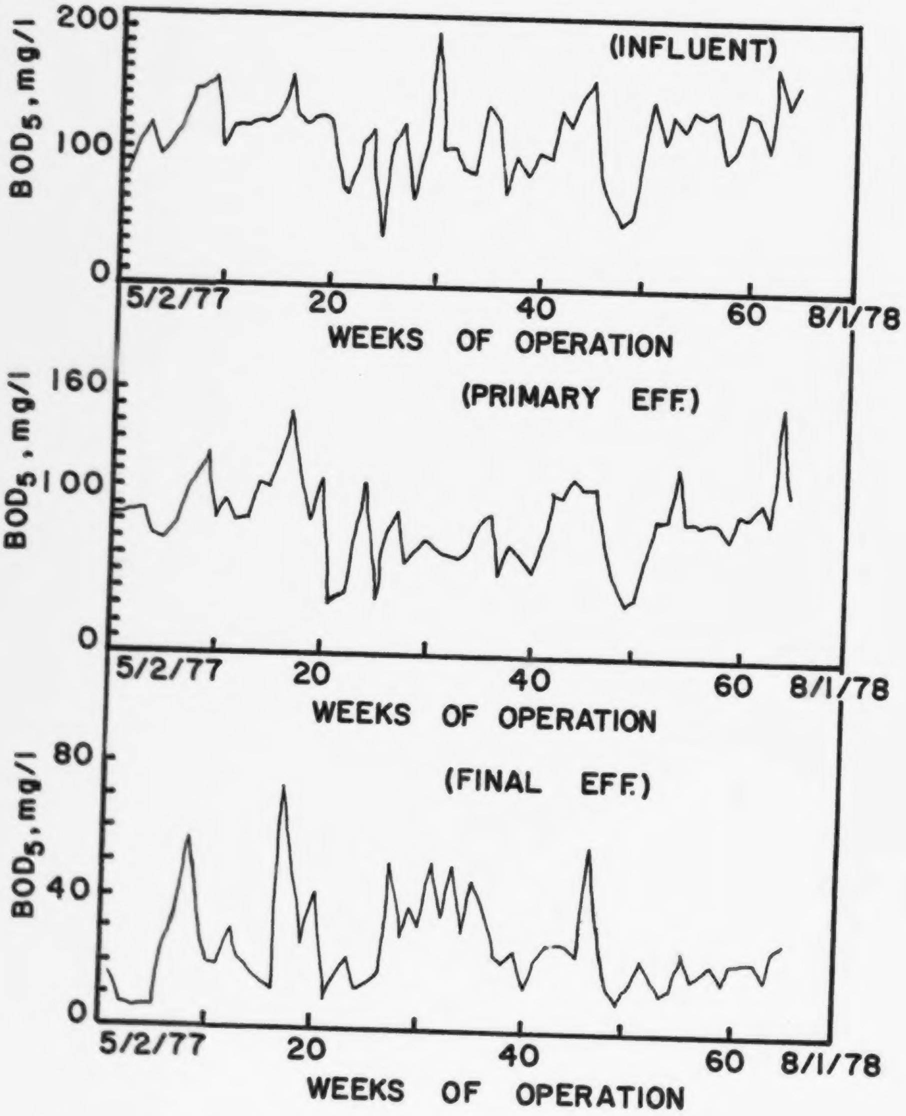


Figure 10. BOD$_5$ data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978.

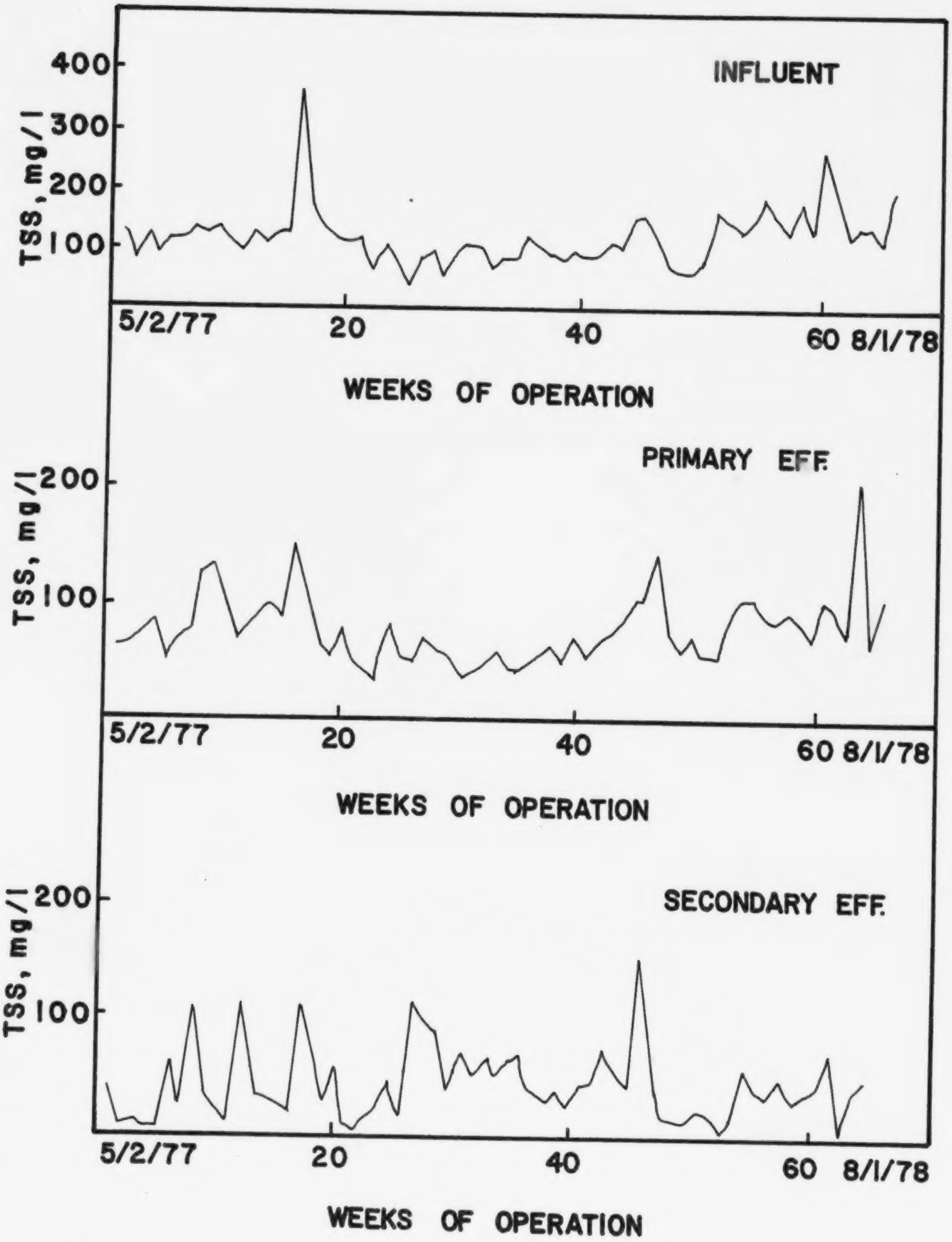


Figure 11. TSS data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978.

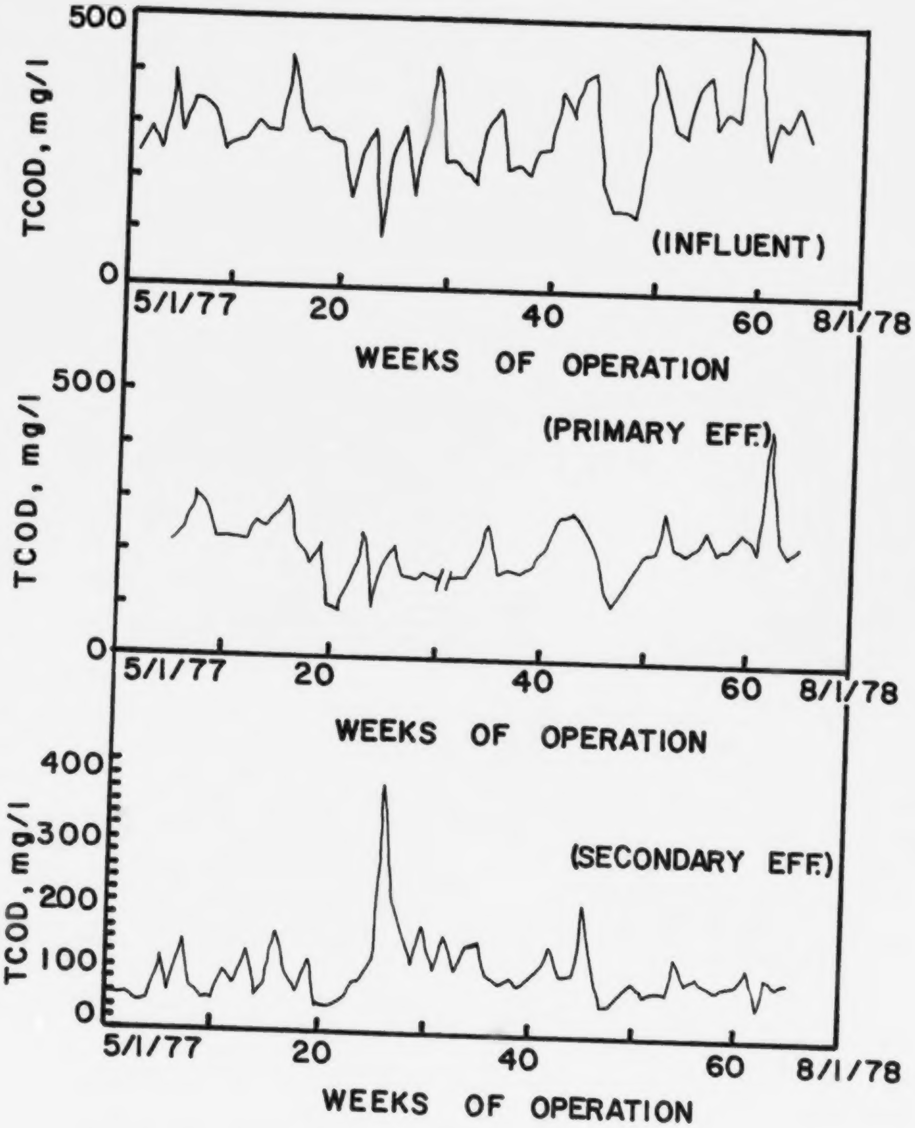


Figure 12. TCOD data for the Binghamton-Johnson City STP from May 2, 1977 to August 1, 1978.

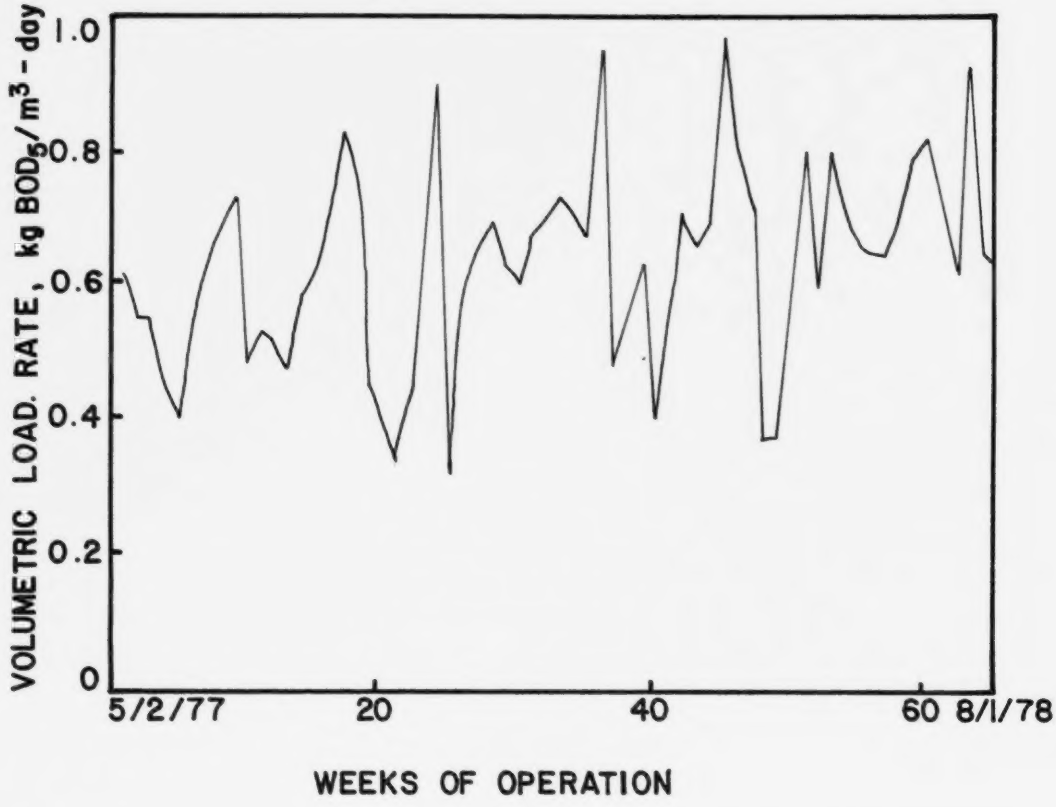


Figure 13. Volumetric loading rate, kg $BOD_5/m^3$-day, for the Binghamton-Johnson City STP aeration basins during the experimental test period.

## TABLE 7. BINGHAMTON-JOHNSON CITY STP PERFORMANCE SUMMARY FOR EXPERIMENTAL TEST PERIOD

| | |
|---|---|
| **Primary Treatment** | |
| Removal Efficiencies: | TSS = 35% |
| | $BOD_5$ = 27% |
| | TCOD = 27% |
| Raw Sludge Flow: | 3,028 $m^3$/day |
| **Secondary Treatment** | |
| Removal Efficiencies: | TSS = 63% |
| | $BOD_5$ = 78% |
| | TCOD = 67% |
| **Gravity Thickeners** | |
| Flow to Anaerobic Digesters: | 250 $m^3$/day |
| Concentration: | 45 gm/ℓ Total Solids |
| **Anaerobic Digesters** | |
| Treatment Efficiencies: | 35% Total Volatile Solids |
| Gas Production: | 3,085 $m^3$/day |

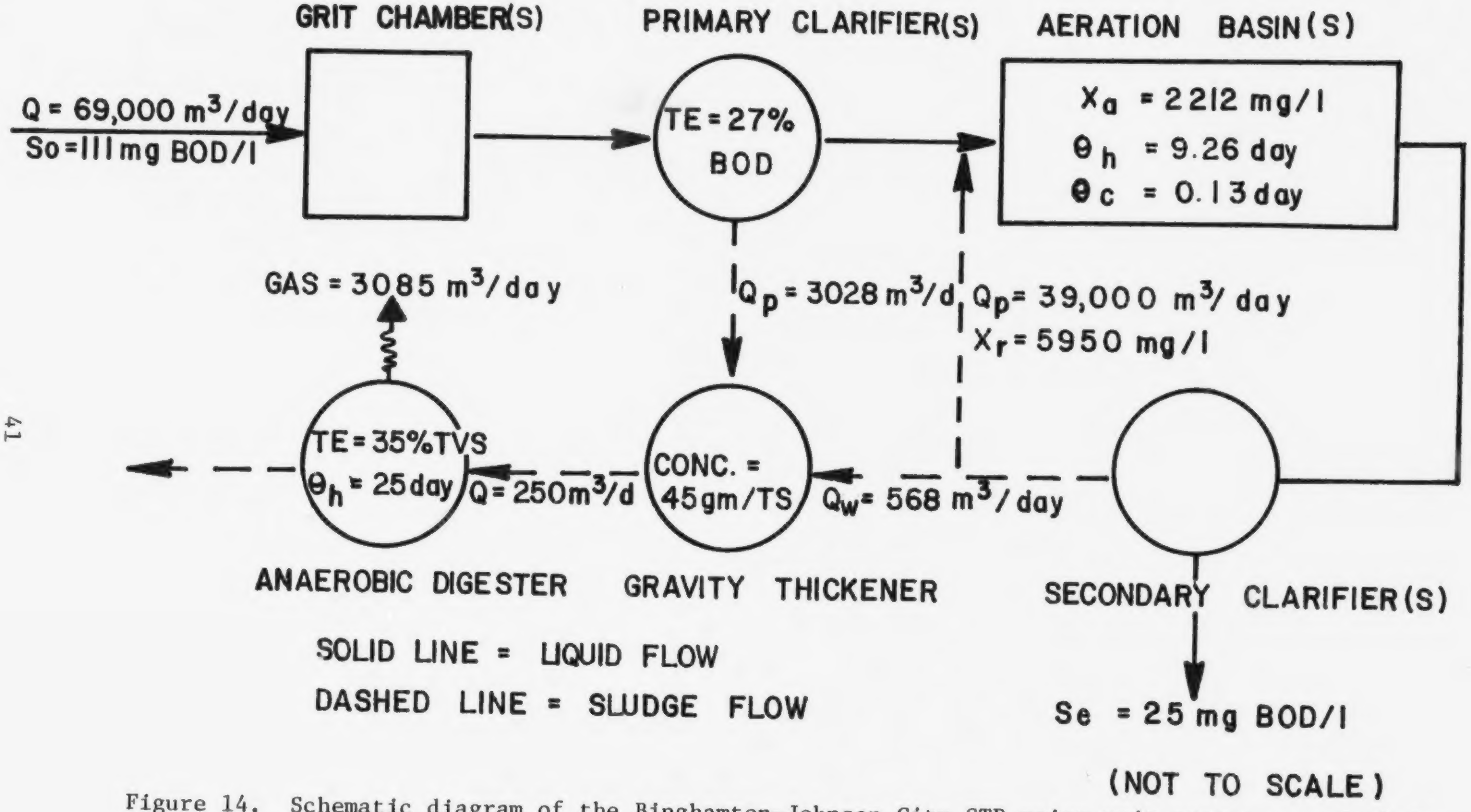

GRIT CHAMBER(S)
PRIMARY CLARIFIER(S)
AERATION BASIN(S)
Q = 69,000 m³/day
So = 111 mg BOD/l
TE = 27% BOD
Xa = 2212 mg/l
θh = 9.26 day
θc = 0.13 day
GAS = 3085 m³/day
Qp = 3028 m³/d
Qp = 39,000 m³/day
Xr = 5950 mg/l
TE = 35% TVS
θh = 25 day
Q = 250 m³/d
CONC. = 45 gm/TS
Qw = 568 m³/day
ANAEROBIC DIGESTER
GRAVITY THICKENER
SECONDARY CLARIFIER(S)
SOLID LINE = LIQUID FLOW
DASHED LINE = SLUDGE FLOW
Se = 25 mg BOD/l
(NOT TO SCALE)

Figure 14. Schematic diagram of the Binghamton-Johnson City STP major unit processes and their respective average performances.

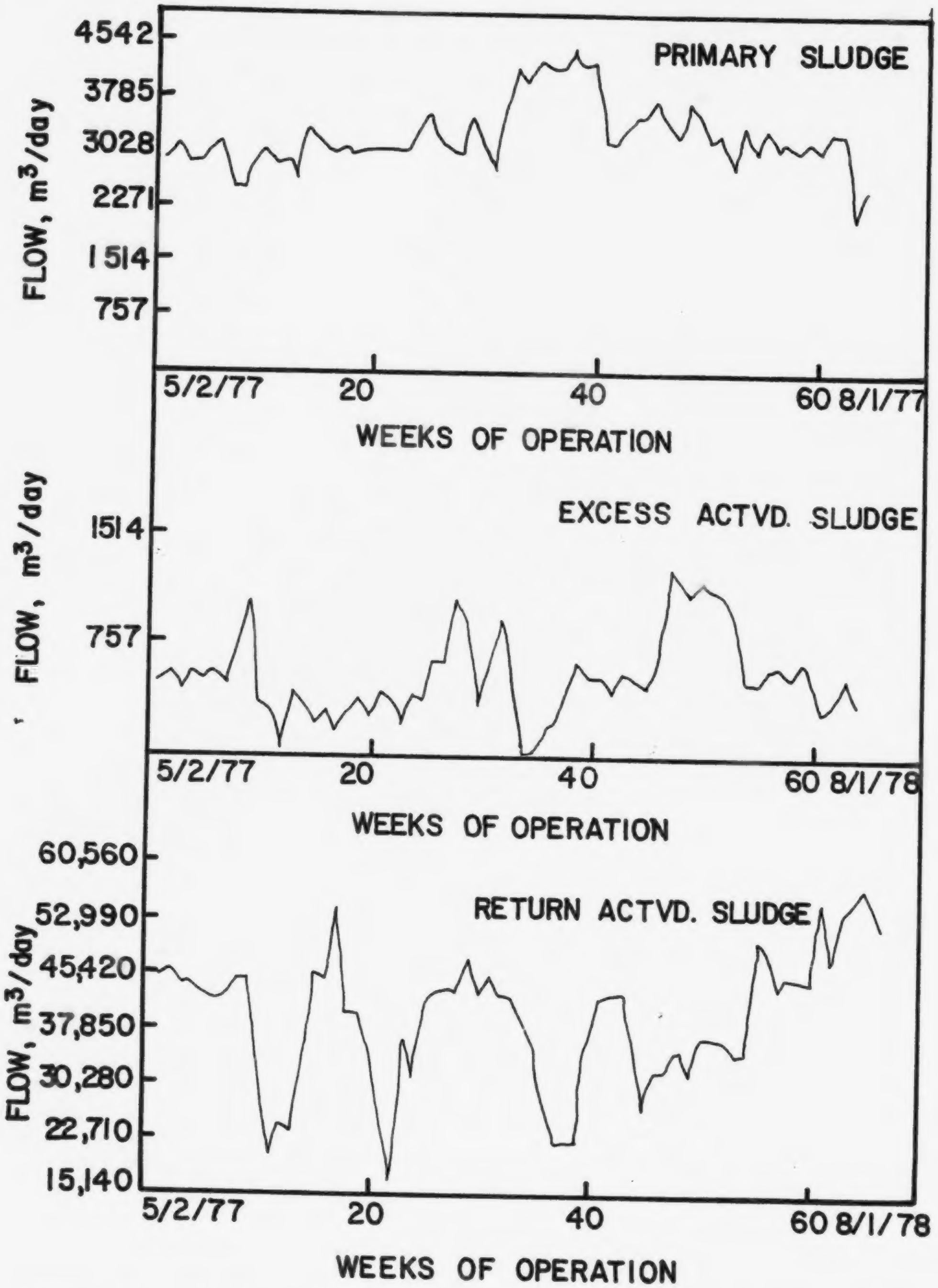


Figure 15. Sludge flows for raw, return, and excess sludges during the experimental test period.

period of this study. It can be seen from these figures that the volumetric ratio of primary to secondary sludge was variable ranging from 3.31 to 10.8 with an average value of 7.1. Increased organic loadings to the treatment plant probably were responsible for increased sludge production, hence, increased excess sludge, since the MLSS was kept low.

Although the characterization of the primary sludge was not taken beyond a flow and concentration basis, due to the limitations of the available data (from the STP records), it was possible to obtain an idea of the quality of the sludge from the secondary system by looking at several parameters. Table 8 presents a detailed summary of the various operational parameters for the secondary treatment system. Note that the sludge age ($\Theta_c$), MLSS, food to microorganism ratio (F/M), and hydraulic retention time ($\Theta_h$) are in line with those suggested for non-nitrifying, complete-mix, or contact stabilization activated sludge process operation (75). Figure 16 shows the MLSS concentrations maintained in the aeration basins as well as the return sludge concentrations during the study. The ratios of sludge returned to total sludge flow (secondary), i.e., $Q_R/Q$, were quite high, as seen in Figure 15, due to the inadequacies of the secondary clarifiers. The average return sludge concentration was only 6,000 mg/ℓ SS, therefore, a large recycle ratio was needed to maintain a minimum $X_a$. It is useful to note here that the secondary sludge was relatively constant with regard to settling ability as measured by sludge volume index (SVI). Figure 17 shows the SVI values over the period of study. The adequate settling ability of the mixed liquor as reflected by the SVI further demonstrates the limited capabilities of the secondary clarifiers.

TABLE 8. DETAILED SUMMARY OF AVERAGE OPERATIONAL PARAMETERS FOR THE ACTIVATED SLUDGE UNIT(S) AT THE BINGHAMTON-JOHNSON CITY STP

| | |
|---|---|
| Hydraulic Retention Time, Aeration Basins, hr.: | 3.17 (± 2.2) |
| Volumetric Loading to Aeration Basins, kg $BOD_5/m^3$-day: | 0.63 (± 0.15) |
| Mixed Liquor Suspended Solids, mg/ℓ: | 2,212 (± 882) |
| Return Sludge Suspended Solids, mg/ℓ: | 5,950 (± 1,730) |
| Sludge Recycle Ratio, $Q_R/Q$: | 0.99 (± 0.01) |
| Food to Microorganism (F/M) Ratio, kg $BOD_5$ applied/kg MLVSS-day: | 0.28 (0.08-1.95) |
| Sludge Age, $\Theta_c$ (mean cell residence time) day: | 9.26 (2.11-62.6) |

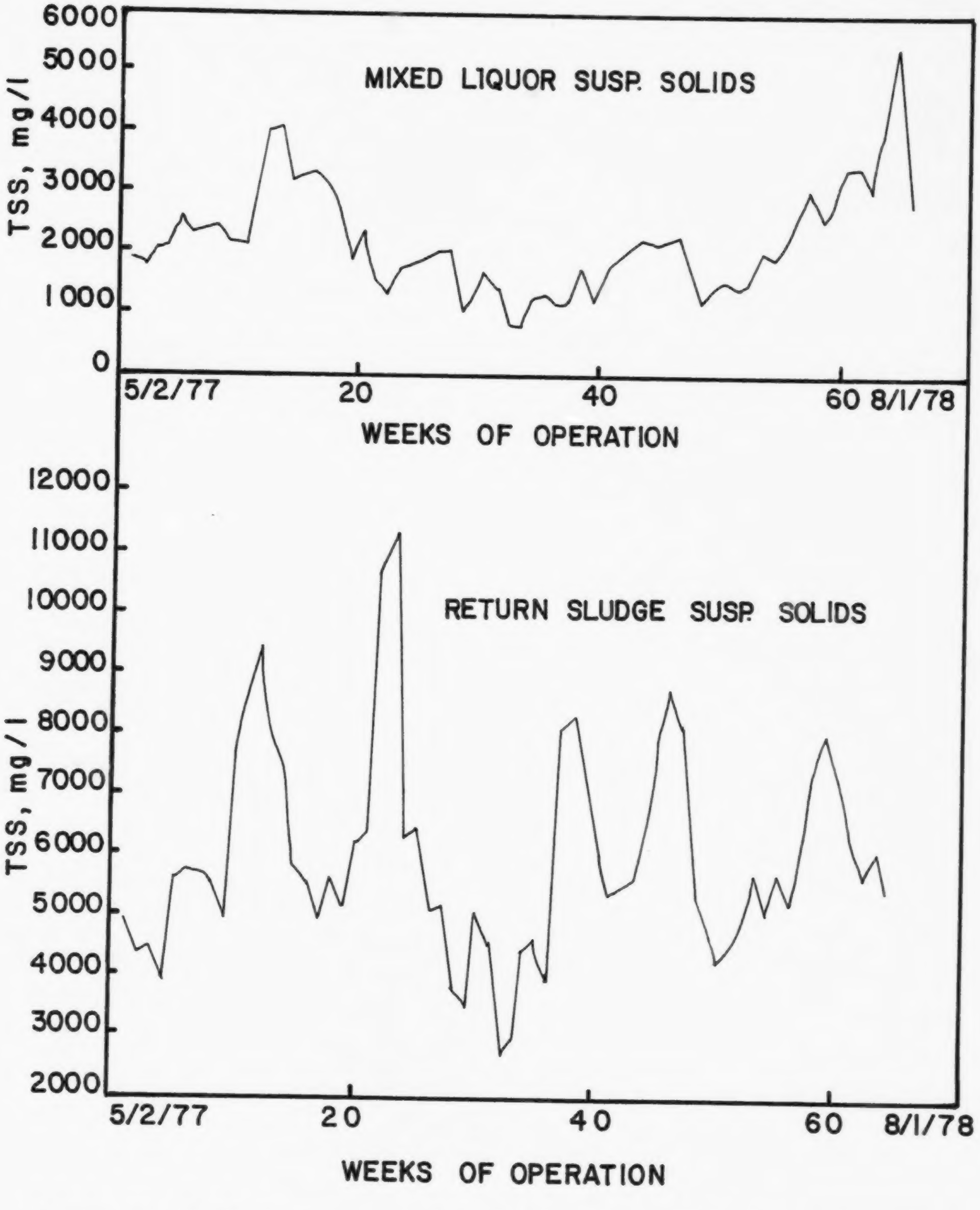


Figure 16. MLSS and return sludge SS concentrations of the Binghamton-Johnson City STP for the period of the full scale study.

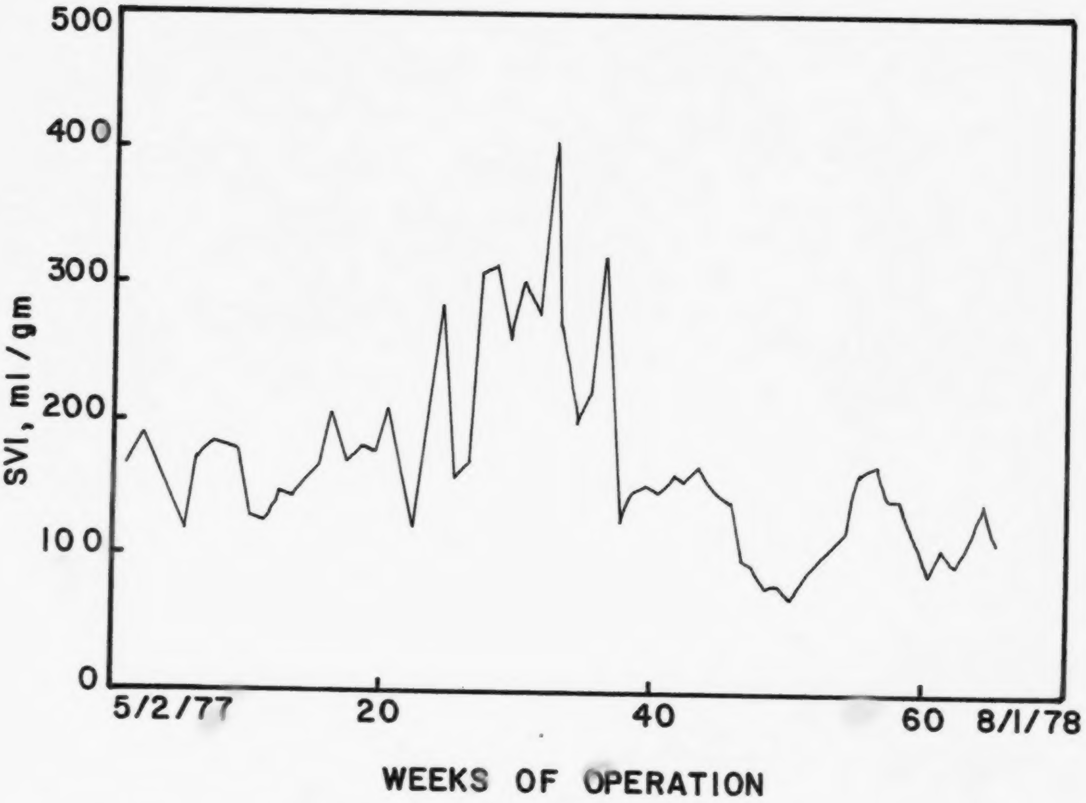


Figure 17. Settling property, as measured by SVI, observed for the activated sludge mixed liquor during the experimental test period.

The combined primary and secondary sludge was gravity thickened, with elutriation used to enhance settling and control septicity for long term sludge storage in the thickeners. The average thickener underflow concentration (see Figure 18) was 4.5% total solids (TS) with a range of 1.5% TS to 7% TS. (For a detailed summary of the thickener effluent composition, i.e., feed sludge for the aerobic (thermophilic) and anaerobic digesters, see Appendix D.) Note that despite fluctuations in the TS concentration of the sludge, the total chemical oxygen demand (TCOD) to total solids (TS) ratio was approximately one throughout the study, thus giving some indication of the autoheating potential of the sludge (see Theoretical Considerations).

From the analyses of the Binghamton-Johnson City STP records, the study by Calcerinos and Spina, and from actual operation of the autoheated aerobic thermophilic digester, the following several general conclusions could be drawn concerning the Binghamton-Johnson City sludge as it came from the thickeners.

(1) The sludge was consistent in volatility, with 65% of the total solids being volatile.

(2) TCOD:TS was approximately one at all times.

(3) Overall sludge age was relatively young due to the high surface overflow rate and limited storage in the primary clarifiers, and to the low $\Theta_c$ in the secondary system.

(4) Sludge appeared readily biodegradable (as reflected in biodegradability studies), despite fluctuations in the flow, loading, and plant performance, with the exception of extremely high flows (caused by precipitation) which caused solids washout or complete malfunction of the unit process(es).

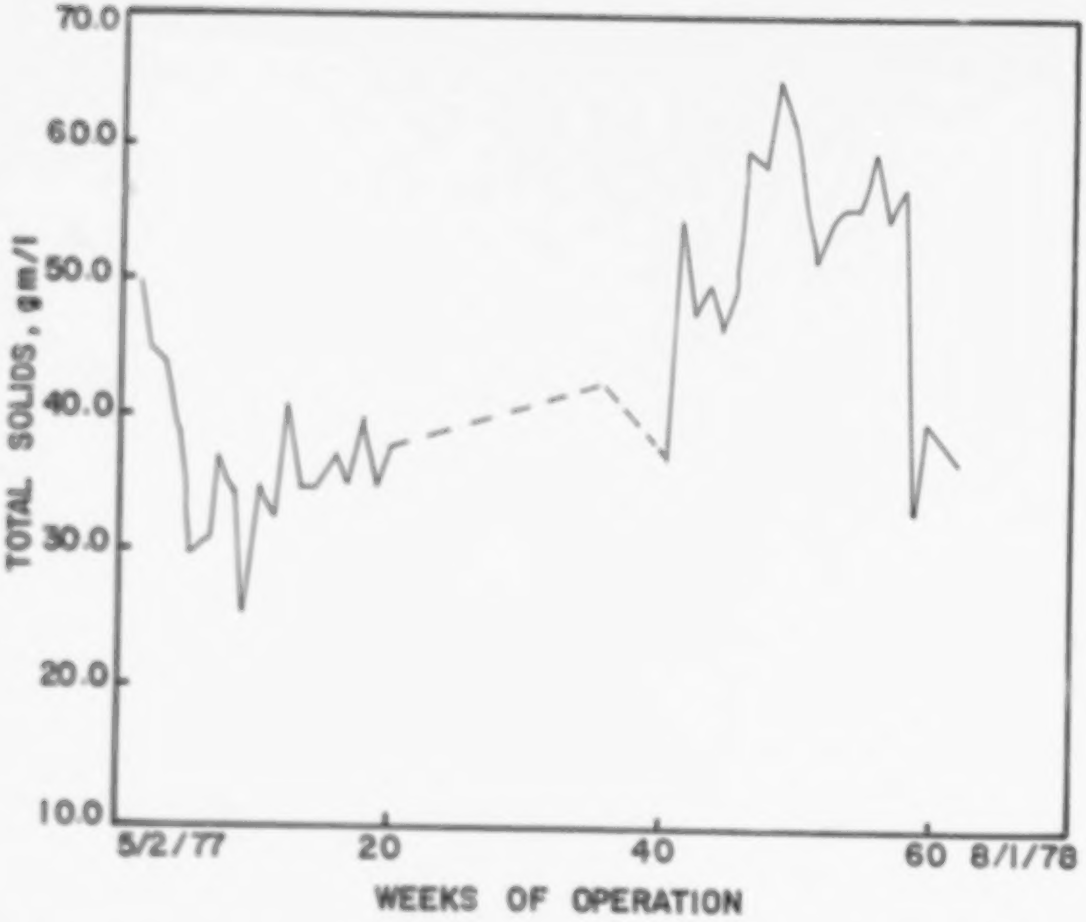


Figure 18. Concentration of TS from the thickener cone(s) during the experimental test period.

## SYSTEM DESIGN AND CONSTRUCTION

### Introduction

Initial investigations of the autoheated aerobic thermophilic digestion process utilizing agricultural wastes were performed at Cornell University by Koenig (76) and Wright (19) with bench scale and pilot scale reactors. Subsequently, a commercially available full scale digestion system (DeLaval, Inc.) was installed at Cornell University's Animal Science Teaching and Research Center (ASTARC), where the system's performance was studied over a two-year period using a dairy waste substrate (6,7). Because of this successful demonstration of the autoheating concept with a dairy waste substrate, it was thought that a similar system could be used to treat municipal sludges. Thus, a full scale system was designed and constructed with supporting funds solicited from the United States EPA at the closest municipality which offered a combination of thickened primary and secondary (waste activated) sludge as well as professional cooperation with Cornell University. The facility of choice was the Binghamton-Johnson City Sewage Treatment Plant located in Broome County, New York, approximately 50 miles from Cornell University, at which this study took place over a period of two years (see previous section). Bench scale digesters, both continuous and batch operated, were designed, constructed, and operated in parallel with the full scale system at the two support facilities (ASTARC and AWML). The ensuing section details the design, construction, and components of both the full scale and bench scale systems.

### Full Scale

The design of the Binghamton-Johnson City system was based upon the digestion facility located at ASTARC and marketed in this country by DeLaval, Inc. The system was designed to enable measurement of mass balances for flow and energy by providing large completely mixed and insulated feed and effluent tanks, each one capable of storing a volume approximately equal to one hydraulic retention period (equal to operating volume of the reactor). Some constraints were placed on the design and construction of the new system because funding was limited and the facility had to be of a temporary nature.

Actual construction of the full scale system was initiated with the delivery of the reactor vessel (fabricated to specifications) and the two tanks (located and leased from a Vestal, N.Y. firm) that would serve as influent (holding) and effluent (collection) tanks. With these tanks and the system's major super-structures in place, the research team was able to assemble the many pieces of equipment and material needed to complete the new full scale aerobic digestion system. The list of construction tasks in Table 9 provides a synopsis of the construction phase (first year) of the project as well as a summary of the system modification and winterization that took place in the second year of study. Figures 19 through 29 demonstrate pictorially the construction progress and the physical location of the project at the Binghamton-Johnson City STP.

A detailed description of the components of the full scale system follows, with a schematic diagram of the system presented in Figures 30 and 31:

## TABLE 9. SUMMARY OF FULL SCALE CONSTRUCTION IN ORDER OF SUCCESSION

### First Year

1. Placement of reactor on tank pad made of railroad ties.
2. Excavation of effluent tank site and placement of said tank.
3. Fabrication and placement of influent and effluent tank lid platforms.
4. Placement of stairway and between-tank walkways.
5. Fabrication and placement of wood and metal safety railings.
6. Fabrication and placement of 15.2 cm sludge delivery pipe from existing thickener pumps to influent tank.
7. Fabrication and placement of mixer (Lightnin' Model 70Q2) and pump (Crisafulli Model CPO2) support mount beams on top of influent tank.
8. Mixer (as above) and mixer shaft-turbine assembled and mounted.
9. Feed pump (Crisafulli Model CPO2) mounted.
10. Placement of sealable sample ports in all tanks (located on lids).
11. Sealing and painting of reactor interior.
12. Sandblasting and painting of influent and effluent tank exteriors.
13. Fabrication and placement of electrical control panel.
14. Delivery power line (208 volt, 3 phase, 60 amp service) from power panel in thickener building to electrical panel on reactor.
15. Aerator and foam cutter assembled and placed.
16. Fabrication and placement of gravity overflow system from reactor to effluent tank.
17. Placement of bottom drain systems in all tanks.
18. Insulation of reactor.
19. Testing and calibration of equipment.

### Second Year

1. Design and placement of 0.6 m x 0.6 m concrete radius in the reactor floor for enhancement of mixing and oxygen transfer.
2. Insulation of influent and effluent tanks.
3. Placement of resistor type heating wires on all exposed valves and pipes (thermostatically controlled).
4. Insulation of all valves and pipes.
5. Placement of pump storage and sampling shed.
6. Installation of effluent pump (Hydromatic Model 30MPV, 240 volt, 3 phase, 1,750 rpm, 7.46 kW).
7. Overhaul of feed pump (see #9 above), i.e., new bearings and impeller.

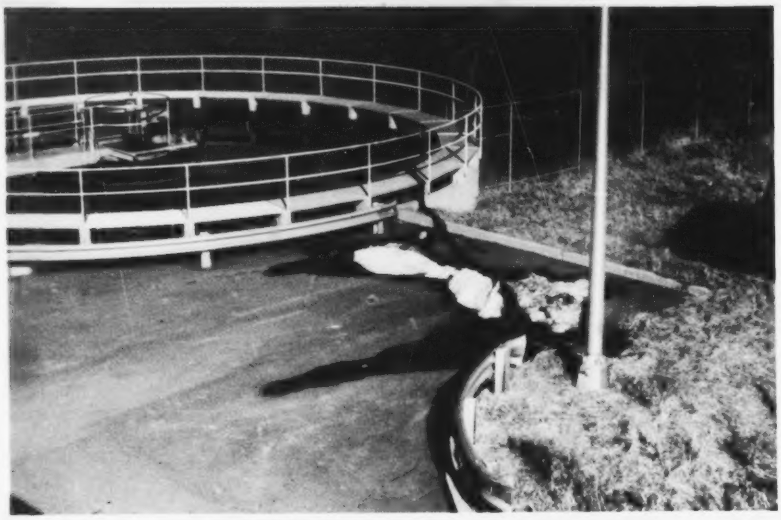


Figure 19: Site of the thermophilic digester test unit location at the Binghamton-Johnson City sewage treatment facility. The unit in the background is one of the sludge thickeners from which sludge was obtained for the study.

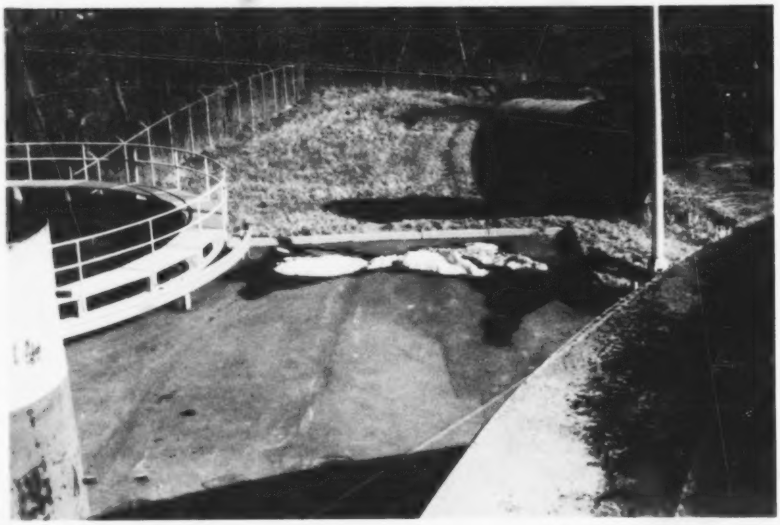


Figure 20: Site of the thermophilic digester study at the Binghamton-Johnson City sewage treatment facility showing the effluent storage tank prior to installation.

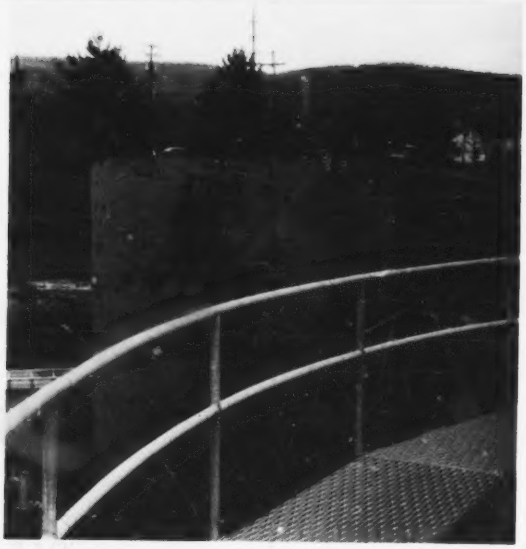


Figure 21: Thermophilic digestion study effluent tank during installation at the Binghamton-Johnson City sewage treatment facility.

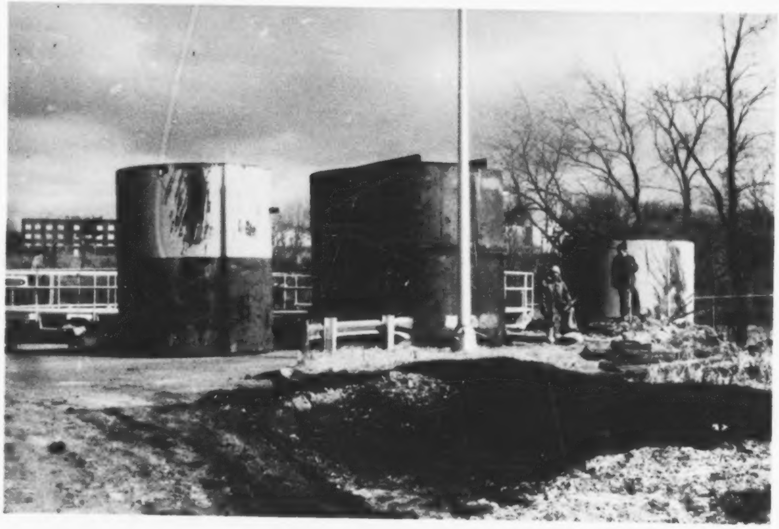


Figure 22: Thermophilic digestion system at the Binghamton-Johnson City sewage treatment facility prior to insulation and piping addition.

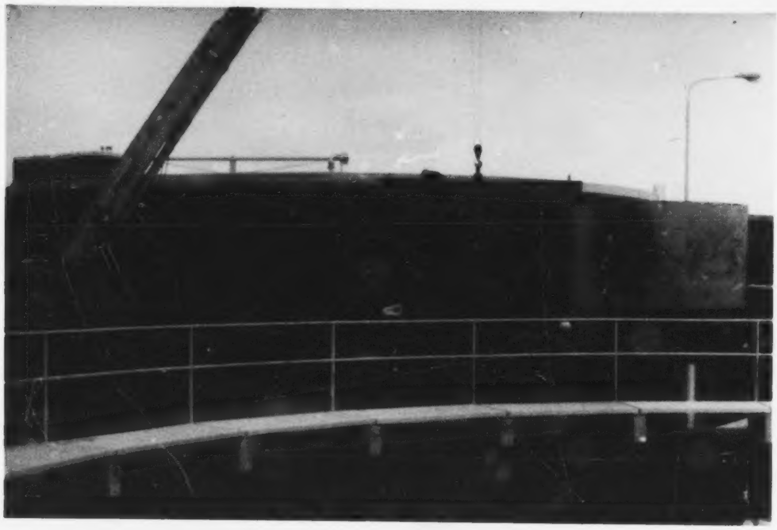


Figure 23: Setting the influent feed storage tank for the thermophilic digestion facility at the Binghamton-Johnson City sewage treatment facility.

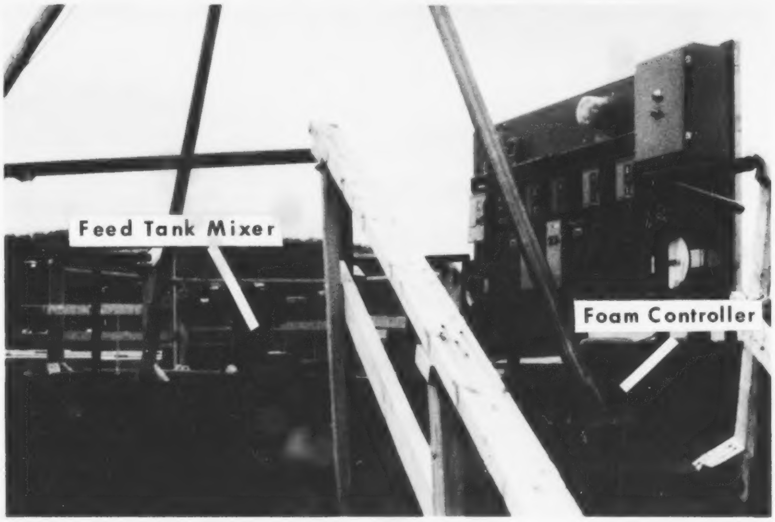

Feed Tank Mixer
Foam Controller

Figure 24: Electrical control panel fabricated for the thermophilic digestion facility.

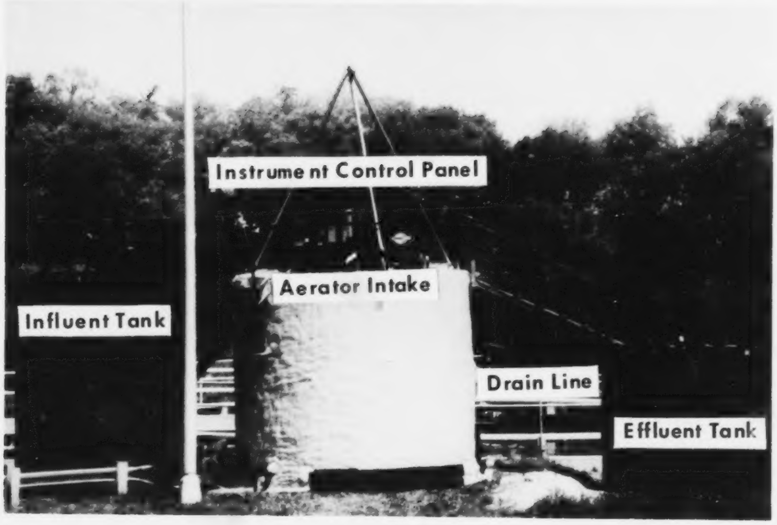

Instrument Control Panel
Aerator Intake
Influent Tank
Drain Line
Effluent Tank

Figure 25: Aerobic thermophilic test facility.

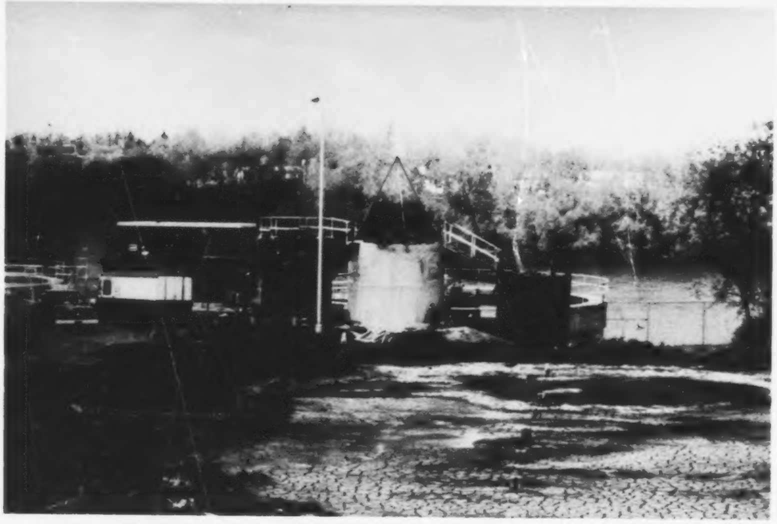


Figure 26: Aerobic thermophilic test facility located beside the sludge thickeners at the Binghamton-Johnson City sewage facility.

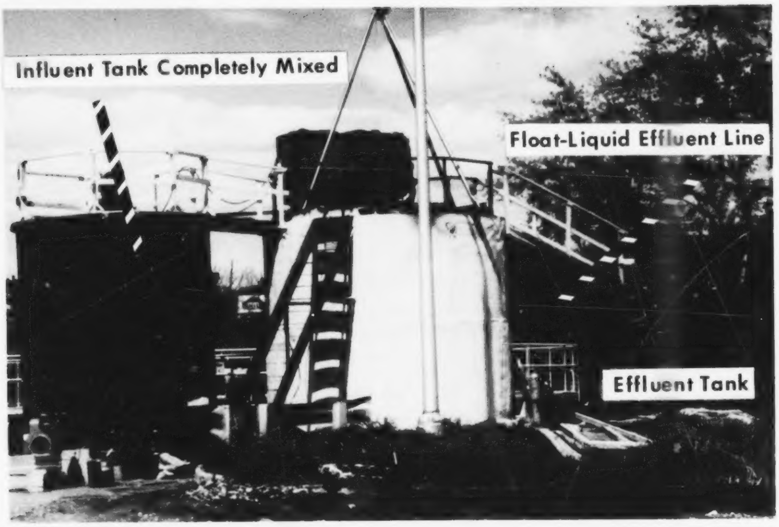

Influent Tank Completely Mixed
Float-Liquid Effluent Line
Effluent Tank

Figure 27: Side view of the completed 28.4 $m^3$ insulated thermophilic reactor, feed tank and effluent tank.

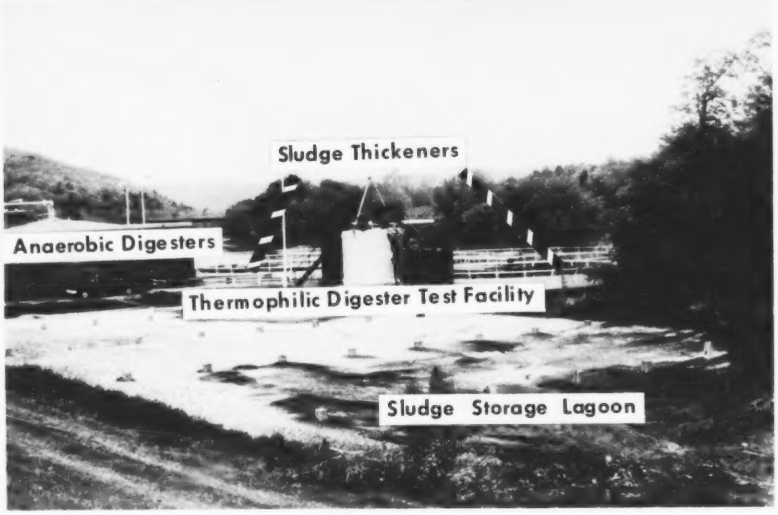

Sludge Thickeners
Anaerobic Digesters
Thermophilic Digester Test Facility
Sludge Storage Lagoon

Figure 28: Relationship of the thermophilic digestion test facility to other unit processes at the Binghamton-Johnson City sewage treatment facility.

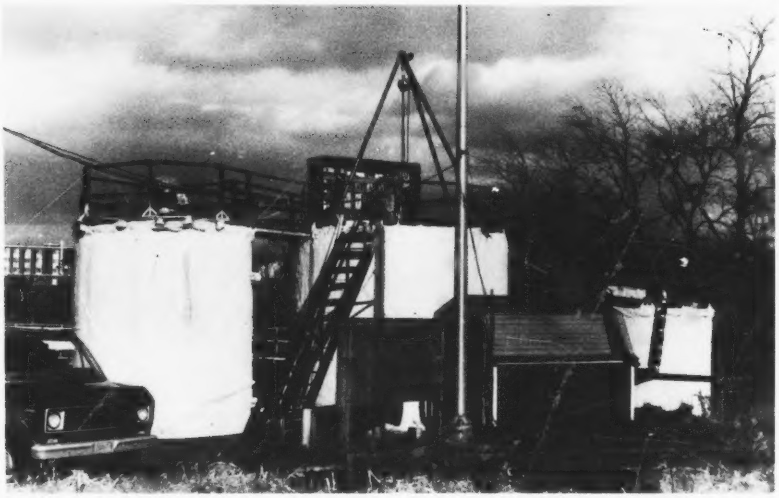


Figure 29: Completed thermophilic digestion system as it appeared in the second year of the study.

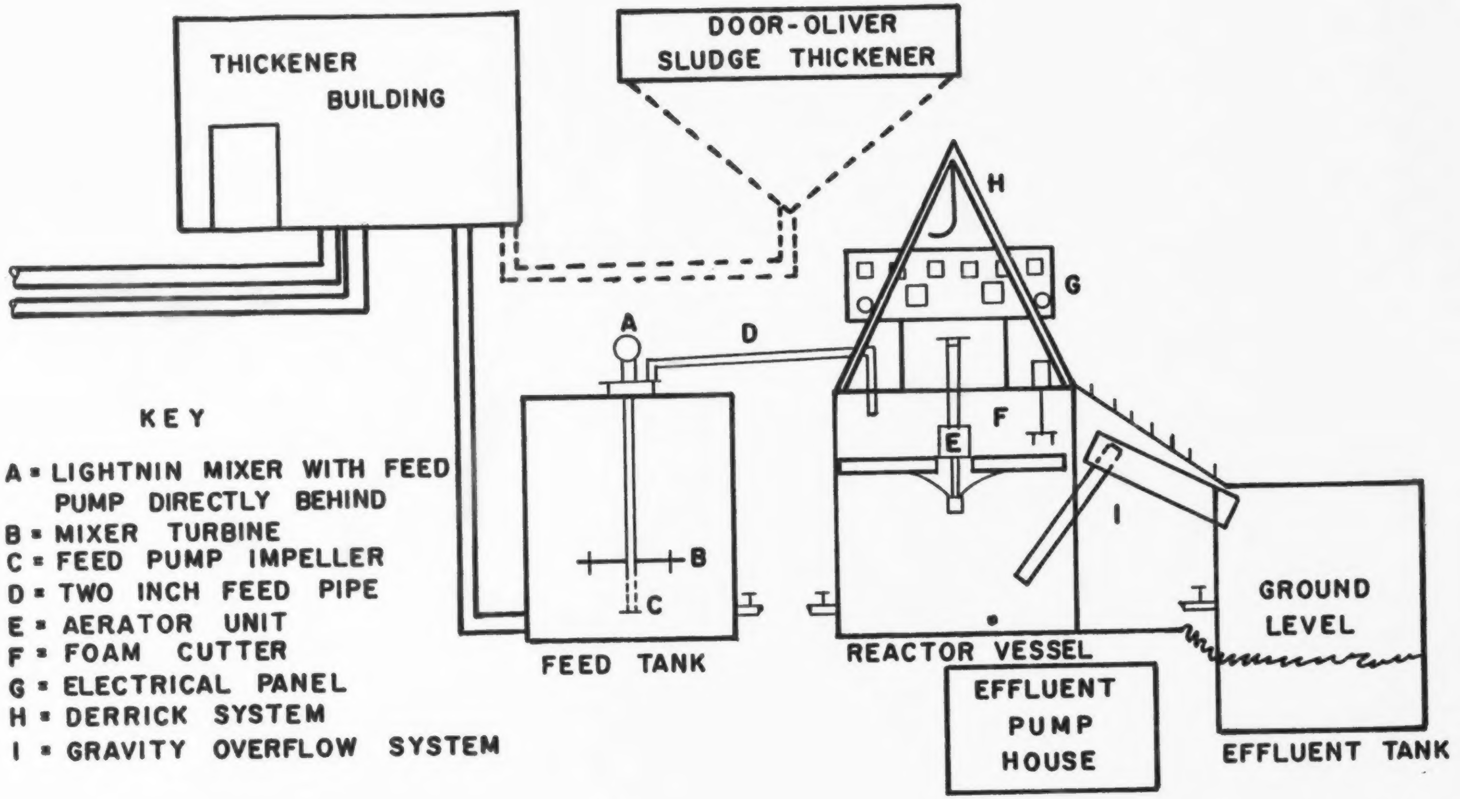

THICKENER BUILDING
DOOR-OLIVER SLUDGE THICKENER
H
G
A
D
F
E
B
C
I
FEED TANK
REACTOR VESSEL
EFFLUENT PUMP HOUSE
GROUND LEVEL
EFFLUENT TANK
KEY
A = LIGHTNIN MIXER WITH FEED PUMP DIRECTLY BEHIND
B = MIXER TURBINE
C = FEED PUMP IMPELLER
D = TWO INCH FEED PIPE
E = AERATOR UNIT
F = FOAM CUTTER
G = ELECTRICAL PANEL
H = DERRICK SYSTEM
I = GRAVITY OVERFLOW SYSTEM

Figure 30. Binghamton-Johnson City full scale thermophilic digestion facility.

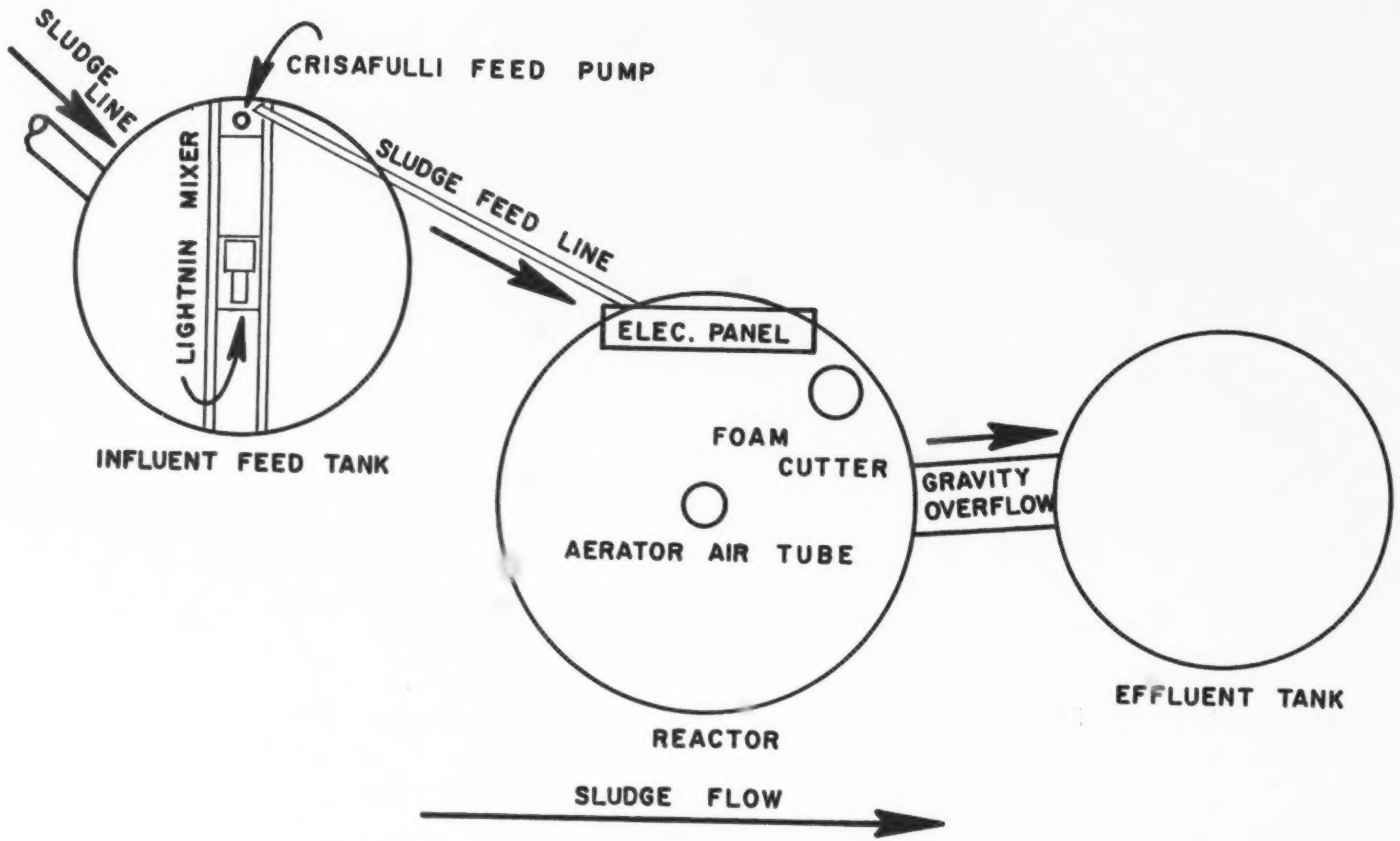
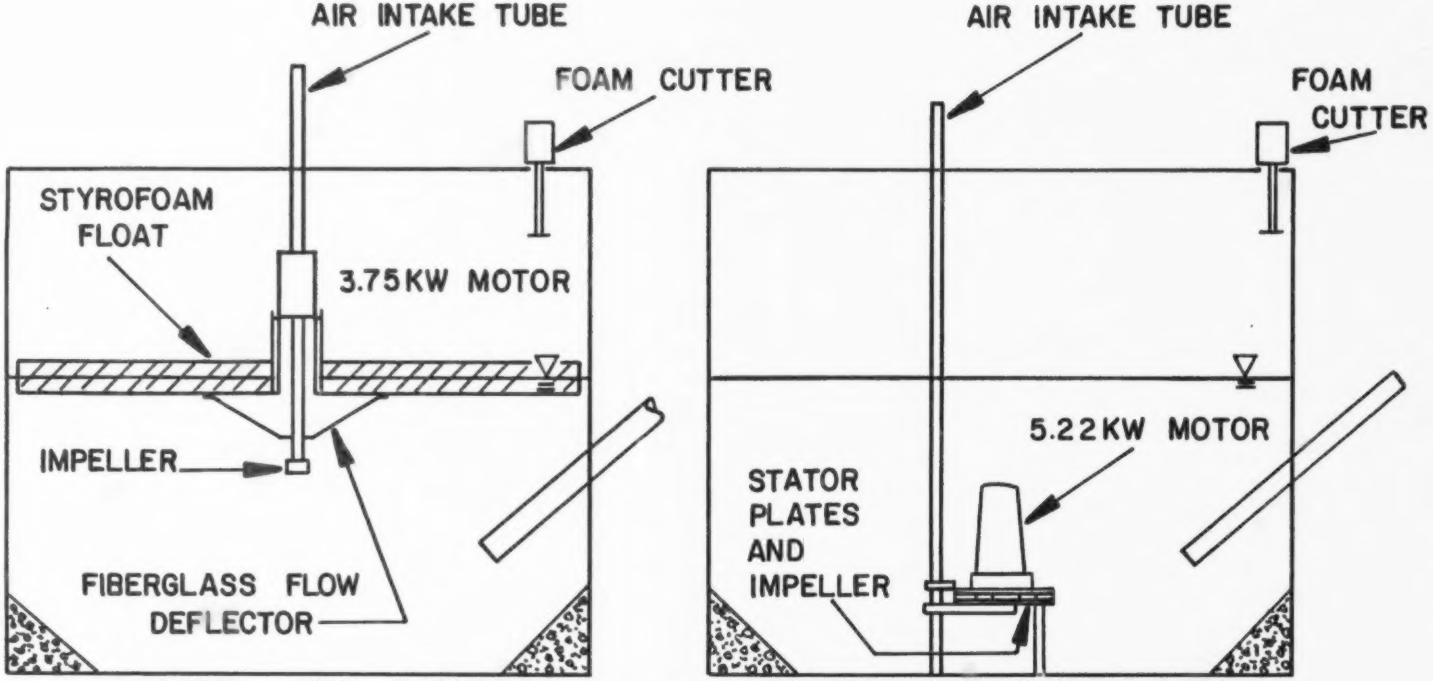

SLUDGE LINE
CRISAFULLI FEED PUMP
LIGHTNIN MIXER
SLUDGE FEED LINE
INFLUENT FEED TANK
ELEC. PANEL
FOAM CUTTER
AERATOR AIR TUBE
GRAVITY OVERFLOW
REACTOR
EFFLUENT TANK
SLUDGE FLOW

Figure 31. Aerial view of Binghamton-Johnson City full scale thermophilic digestion facility showing major components.

(1) The experimental site was chosen due to its close proximity to the sludge thickener building from which a 15.2 cm diameter pipe line was constructed for delivery of thickened primary and secondary sludge to the experimental influent tank. This delivery was accomplished with existing positive displacement sludge pumps located in the basement of the thickener building. The influent storage tank had an effective volume of 28.4 $m^3$ (3.8 m in height x 3.1 m in diameter). Mounted upon this tank (i.e., on 2-20 cm wide structural steel beams set 31 cm apart) was a turbine mixer (Lightnin' 240 volt, 3 phase, 0.746 kW, 36 rpm, maximum viscosity rating equal to 5,000 cps) with an axial flow impeller. The mixer was time controlled to operate for a period of fifteen minutes every two hours with the subsequent agitation providing a homogeneous sludge mixture for feeding to the reactor. A 5 cm diameter, vertical drive, centrifugal pump (Crisafulli Model CP02, 240 volt, 3 phase, 3.75 kW, 1,750 rpm) was also mounted on the influent tank. The pump was used to deliver fresh sludge to the reactor and was timer controlled to operate for a number of minutes each hour, thereby allowing careful control of the solids retention time (SRT). The influent tank had a wooden platform top, and the entire tank was insulated with a 10.2 cm thickness of polyurethane foam.

(2) The reactor was fabricated, by Binghamton Steel and Fabricating Co., of 0.5 cm steel plate, hot-rolled and welded, and fitted with a removable lid and 4-15.2 cm diameter sealable sample ports. To provide the proper geometry for mixing, according to earlier studies (6,7,16,25, 26), the liquid depth in the reactor was set equal to one-half of the reactor diameter, thereby yielding an effective depth of 2.13 meters. To further aid in mixing, the intersection between the floor and walls of the reactor was filled with concrete to form a 0.6 meter by 0.6 meter reactor radius. The necessity of such a radius was determined by Fuchs (25) and further documented by this study (see subsequent section). A freeboard for foam dissipation was provided by an additional 1.54 meters of reactor height above the operating liquid depth. The reactor was totally enclosed and insulated with 10.2 cm of polyurethane foam to provide efficient conservation of heat in the system.

(3) Aeration and mixing in the reactor were accomplished by one of the two aerators tested. During most of the study, the DeLaval Separator Company centri-rator (240 volt, 3 phase, 3.7 kW, 1,750 rpm) was used. This aerator was suspended in the liquid by means of four styrofoam blocks set equidistant from each other. This flotation system insured that the optimum immersion depth, as determined by Cummings and Jewell (6, 7), and established by the manufacturer, was maintained for the impeller. This aerator was of the self-aspirating kind, i.e., the vacuum created at the center of the aerator impeller draws air down the hollow air intake tube and draws liquid up from the bottom center of the tank, thereby providing aeration of the liquid at the impeller. The other aerator studied was an LFE Corporation Midland-Frings (240 volt, 3 phase, 5.2 kW, 1,750 rpm) self-aspirating turbine aerator. This aerator sat on a tripod 30 cm from the reactor bottom with the impeller (turbine) spinning within the stator plates to draw air down the intake tube and move the liquid and air mixture to the reactor perimeter. Note that on the Midland-Frings aerator the impeller was encased within the stator, whereas the DeLaval centri-rator impeller hung free in the liquid. See Figures 32 through 34 for additional details.


AIR INTAKE TUBE
FOAM CUTTER
STYROFOAM FLOAT
3.75 KW MOTOR
IMPELLER
FIBERGLASS FLOW DEFLECTOR
AIR INTAKE TUBE
FOAM CUTTER
5.22 KW MOTOR
STATOR PLATES AND IMPELLER

REACTOR WITH FLOATING CENTRI-RATOR IN PLACE

REACTOR WITH MIDLAND-FRINGS UNIT IN PLACE

Figure 32. Comparison of the two aerators tested in this study and their location in the reactor.

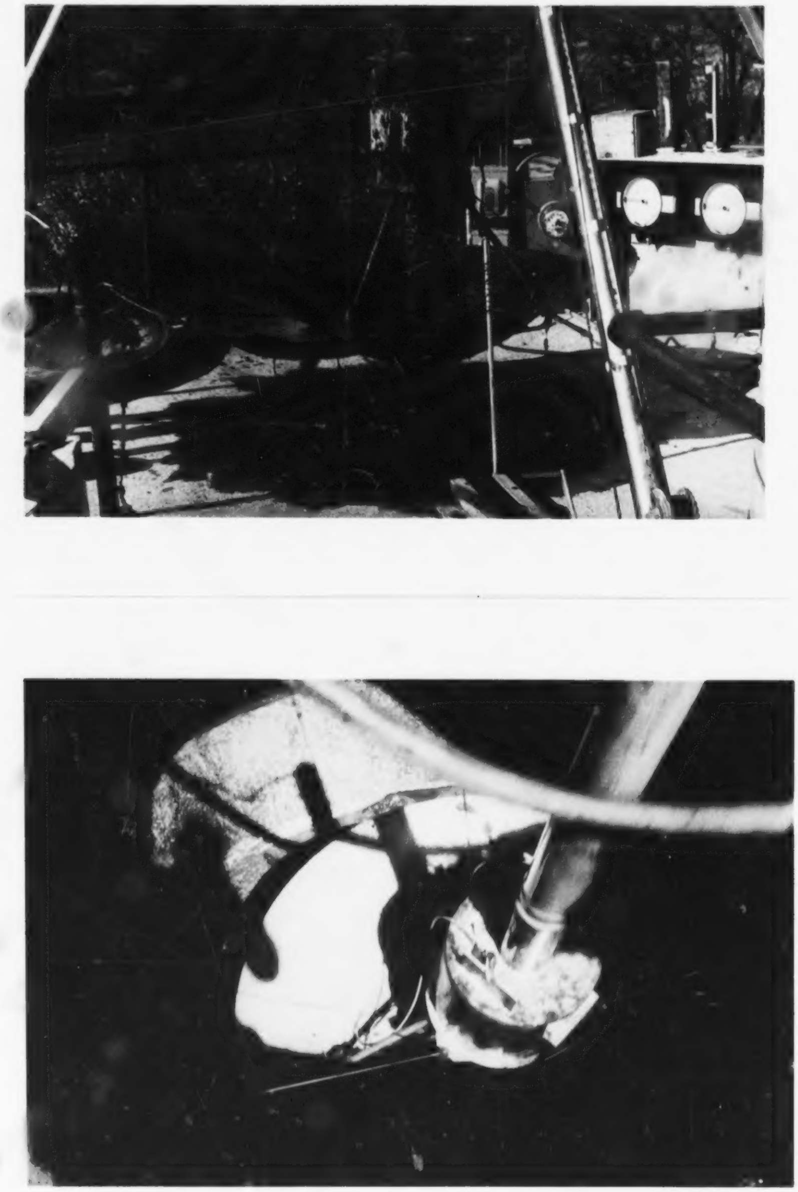


Figure 33: The centri-rator aerator suspended above the reactor (top photograph) and in place in the reactor (bottom photograph).

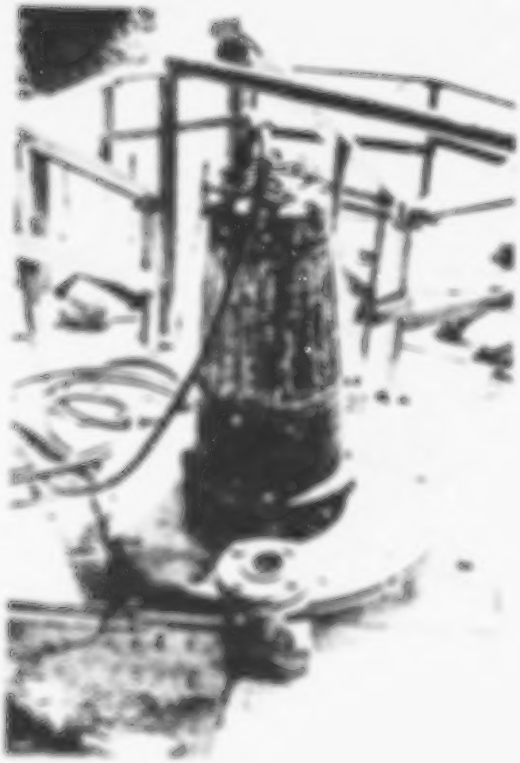


Figure 24: The Midland-Fringe aerator on top of the reactor without support tripod and air intake tube in place.

(4) Control of the foam created biologically and chemically was accomplished with a mechanical foam cutter (DeLaval, 240 volt, 3 phase, 1.5 kW, 1,750 rpm). The unit was equipped with a bladed circular impeller and mounted on rubber vibration mounts near the outer edge of the top of the reactor. Note that the impeller operated 1.0 meter above the liquid surface.

(5) The electrical control panel was centrally located on top of the reactor, with all of the electrical equipment controlled from this panel. The temperature recorders (Partlow Model RFHT and Model RFHTT, recording seven day thermometers, 120 volt, effective sensitivity range of -35°C to 75°C) for the ambient, feed sludge, and reactor temperatures were mounted within the electrical control panel. The thermal element was the temperature sensing portion of the Partlow temperature recorder and consisted of a hermetically-sealed, mercury-filled assembly with a bulb, capillary tube, and head assembly. The element, when assembled to the recorder, provided mechanism actuation through the expansion and contraction of mercury in the bulb due to temperature changes. Power input to the aerator was monitored with a conventional in-line kilowatt meter located on the electrical control panel. Changes in ambient barometric pressure were monitored with an aneroid barometer mounted on the control panel.

(6) The effluent tank was identical to the influent tank, with the exception that it was set approximately one and one-half meters into the ground to facilitate gravity overflow from the reactor and thereby provide the required effluent volume needed (28.4 $m^3$). Note that an entire retention period's effluent was collected before it was sampled and analyzed. This enabled collection of samples that were representative of the entire test period and thus minimized variations during the test period. The gravity overflow consisted of a 15 cm black steel pipe, 2.0 meters in length and set at a 60° angle to the reactor with one-half of the pipe length extending from the reactor and connecting to a 30 cm diameter PVC pipe. The PVC pipe acted as a trough to carry liquid and foam from the overflow pipe to the effluent tank. The effluent tank had a wooden platform top and was insulated with a 10.2 cm thickness of polyurethane foam.

(7) Mixing of the effluent tank for sampling purposes and the subsequent evacuation of the tank was accomplished by a self-priming centrifugal pump (Hydromatic Model 30 MPV, 240 volt, 3 phase, 7.46 kW, 1,750 rpm). This pump was housed in a small wooden building located beside the reactor which was insulated and heated to prevent freezing of the pump. The building also served as a sample preparation and equipment storage area. Sludge from the effluent tank was pumped to the nearest sewer interceptor (leach field from the sand drying beds) where it was then combined with the influent sewage and returned to the head of the STP.

Bench Scale

Well defined biological process kinetics are most easily determined from laboratory studies conducted with small reactors. The maximum biodegradability of municipal sewage sludge under aerobic thermophilic conditions, and kinetic description of the microorganism decay coefficient, half velocity constant, maximum rate of substrate utilization, specific utilization rates, growth yield coefficient, and maximum net specific growth rate of the microorganisms are among the parameters of interest. In an effort to define many of these parameters, continuously fed, semi-continuously fed, and batch fed reactors were designed and operated in conjunction with the full scale reactor under similar conditions.

The continuously fed and semi-continuously fed bench reactors were designed to mimic the full scale system in that they were complete-mix, steady-state systems. The organic removal rates, efficiencies, and reaction rates measured with the closely controlled thermophilic bench units could be used to predict and document the full scale results. On the other hand, batch-like reactors were designed to determine the ultimate biodegradability of the sewage sludge, since several investigators (2,6,7,11, 19,21,76) have shown that the autothermal process is dependent upon the biodegradable substrate available. Information on dewaterability, settleability, and the fate of heavy metals could also be gathered with the long term batch studies. The following section characterizes the bench scale systems' optimized (see results section) design and construction.

General--
Each of the biological reactors was immersed in a hot water bath kept at a constant temperature of 65°C to maintain a constant 50°C within the reactors. Four water baths were fabricated from bovine feed bins (measuring 70 cm x 40 cm x 40 cm) and insulated with styrofoam sheets (5 cm thickness). The water in each bath was heated by a portable constant temperature circulator (Haake Model E52, 120 volt) attached to the edge of the water bath. A one-layer thickness of small plastic balls (one cm diameter) was floated on the water to prevent evaporation and subsequent overloading of the heaters due to excessive heat loss through evaporation.

Each reactor was provided with a separately controlled diffused aeration system. Air was supplied to the diffusers via an air compressor (Worthington Model 5EE8). Air pressure to the regulator was maintained at 30 p.s.i. and was distributed by several manifolds to the reactors. Airflow to the reactors was controlled by airflow meters (Dwyer Precision flow meters) and dispersed within the reactor by porous sandstone diffusers. The excessive foam produced in the aeration of the sewage sludge was controlled by laboratory stirrers (Eastern Industries Model 5VB, 0.043 kW, 120 volt), one for each reactor, equipped with dual blade impellers (2.5 cm x 8.0 cm). The impeller was set at a level 5.0 cm to 20 cm above the liquid level, dependent upon the amount of foam produced.

Continuous and Semi-continuous Reactors--
Two cylindrical plexiglass reactors, each 25 x 61 cm, were designed and built to hold from 15 to 21 liters of sludge, i.e., each reactor had an adjustable liquid level, with the remaining reactor volume used as foam freeboard. Each of these reactors was contained in its own water bath and supplied air as described previously. Figure 33 details the reactors and their components.

The continuously fed unit, i.e., hydraulic retention times (HRT) less than 5 days, was supplied sludge on a timed interval basis, once per hour, by a positive displacement chemical feed pump (Precision Products Model [illegible]), and sludge was fed to the semi-continuous reactor twice per day by research personnel. Reactor contents were sampled by an effluent line (gravity overflow) that retained the given (design) liquid volume for the continuously fed reactor, thus simulating the full scale system. For both units, feed sludge and effluent sludge were refrigerated at 2°C and both sludges were tested on a regular basis.

To prevent excess evaporation losses from the reactor, the influent air was partially saturated with water by bubbling it through a column of water (65°C). Make-up water (tap water) was added to the reactors to compensate for additional evaporation losses on a daily basis.

Batch Mode Reactors--

A group of reactors (four 30 x 15 cm, and four 23 x 19 cm) were designed and built of plexiglass to hold three liters of sludge, with the subsequent remaining reactor volume used to control foam. Two reactors were placed in one water bath and connected to various support equipment as illustrated in Figure 36. Evaporation losses from the reactor were controlled by partially saturating the influent air with water and by the addition of make-up water (tap water).

The reactors received sludge only at time zero or day zero. The sludge was aerated for 25 to 35 days, with subsequent changes in the physical and chemical composition of the sludge monitored on a routine basis, i.e., actual testing of the sludge was once per week during operation.

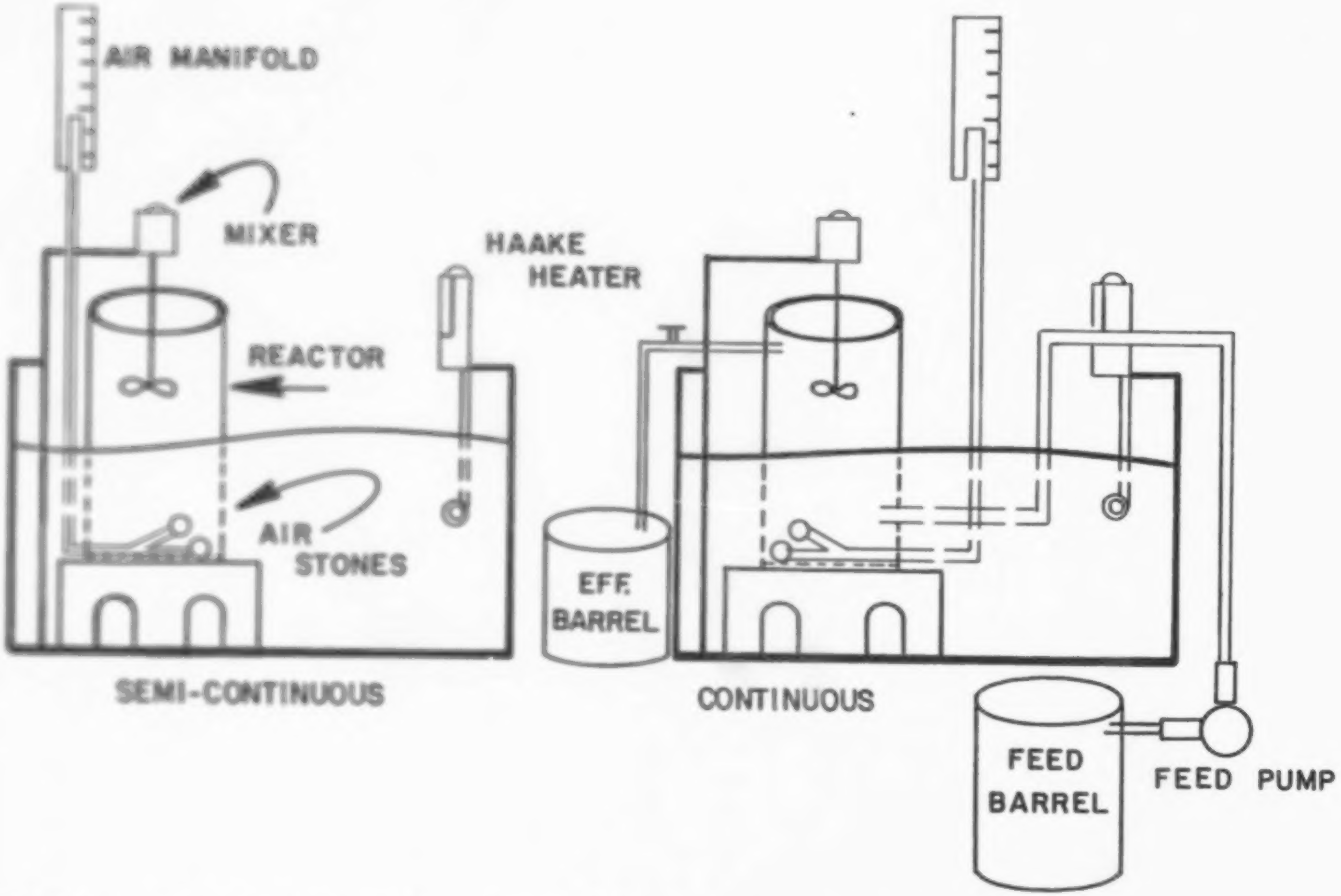

AIR MANIFOLD
MIXER
REACTOR
AIR
STONES
SEMI-CONTINUOUS
HAAKE
HEATER
EFF.
BARREL
CONTINUOUS
FEED
BARREL
FEED PUMP

Figure 35. Continuously and semi-continuously fed laboratory scale reactors.

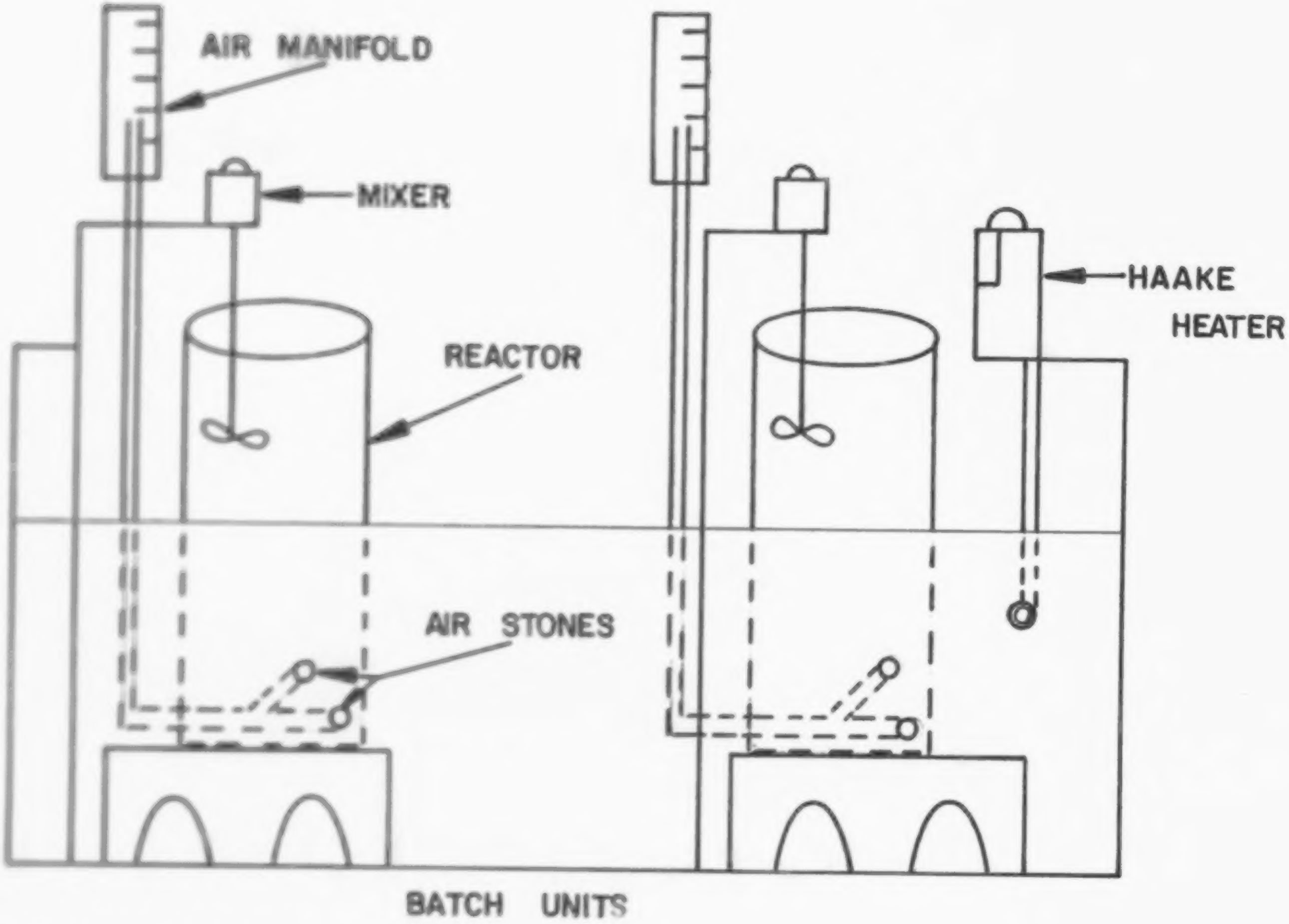

AIR MANIFOLD
MIXER
REACTOR
AIR STONES
HAAKE HEATER
BATCH UNITS

Figure 36. Batch-mode laboratory scale reactor systems.

# SECTION 6

# MATERIALS AND METHODS

## THERMAL AND OXYGEN TRANSFER STUDIES

Studies utilizing clean tap water were conducted to determine the specific heat loss characteristics and aerator (only with the DeLaval centri-rator) performance characteristics of the experimental digester facility. These studies were conducted in the fall of 1977 at the beginning of the second year of the project. Tap water tests were originally scheduled to be performed after completion of construction and prior to start up of the system with sludge; however, equipment delivery and winter weather delays forced the system start up immediately upon completion of construction to prevent any further delays in the testing program.

Following the shut down of the system at the end of the first year, the reactor was emptied and thoroughly cleaned using a steam cleaner. Care was taken to remove all sludge deposits thereby eliminating the possibility of interference with the reaeration studies. Subsequently, modifications to the reactor were performed (see Process Design and Construction) and the reactor was filled with 28.4 $m^3$ of clean tap water at a temperature of 15°C.

In order to study the thermal losses and aerator performance at elevated temperatures, i.e., temperatures exceeding 60°C, the tap water in the reactor had to be heated to approximately 65°C. The following equation allows the calculation of thermal energy needed to raise the temperature of the tap water in the reactor from 15°C to 65°C.

$$Q = (\forall)(T_F - T_O)(D_L)(C_{PL}) \qquad (14)$$

where $Q$ = energy required to raise liquid temperature from $T_O$ to $T_F$, Kcal

$\forall$ = volume of the liquid, $m^3$

$T_O$ = initial temperature of the liquid, °C

$T_F$ = final temperature of the liquid, °C

$D_L$ = density of the liquid, $kg/m^3$

$C_{PL}$ = specific heat capacity of the liquid, Kcal/kg-C.

From this equation (14), it was found that $1.42 \times 10^6$ Kcal were required. A steam generator (Marlsbury Model 250, kerosene fired) with a capacity of $1.89 \times 10^6$ Kcal/hr was located and subsequently connected to the reactor in a manner which allowed injection of the steam into the liquid at 2.0 m below the water surface with water taken from the bottom drain of the reactor and used to continuously fill the boiler on the steam generator. Thus, this closed loop water heating system allowed the tap water to be heated to 65°C in approximately 15 hours.

Upon completion of the heating of the reactor, oxygen and thermal transfer studies were initiated. Various analytical and support equipment were set up as shown in Figure 37. The nonsteady state chemical method (78) was used to determine the oxygen transfer rate of the centri-rator. Sodium sulfite ($NA_2SO_3$) in the presence of a catalyst, cobalt chloride ($CoCl_2$), was used to deplete the dissolved oxygen concentration. A deviation from Standard Methods was made here because it has been shown (78,79,80,81) that high cobalt in concentrations (2 to 5 mg/ℓ) seriously interfere with the Winkler-determined dissolved oxygen concentration. Therefore, cobalt chloride was added to the tap water in the reactor to achieve a concentration of 0.1 to 0.2 mg/ℓ. Both chemicals were premixed in 21 liter containers prior to addition to the reactor tap water.

Aerator performance characteristics were studied at three temperatures, with the first being 60°C. Duplicate runs were performed at each temperature. After the addition of the oxygen depletion chemicals and following the commencement of aeration, the rate of increase in dissolved oxygen concentration was measured with the Winkler method (azide-modification) (77) and with a dissolved oxygen (D.O.) meter (Orbisphere brand). The D.O. meter was standardized against saturated (oxygen) water at test temperature (60°C) and a known atmospheric pressure utilizing the azide-modification of the Winkler method (77). Temperature, reactor and ambient, aeration power (kW), and airflow were monitored simultaneously with the increase in dissolved oxygen (D.O.). Airflow was measured by two methods. The first method utilized a manometer and Bernoulli's principle concerning the pressure differential across an orifice of known dimensions. The air velocity can be calculated using Bernoulli's equation as follows:

$$Z_1 + P_1/\gamma_A + V_1^{\,2}/2g = Z_2 + P_2/\gamma_A + V_2^{\,2}/2g \qquad (15)$$

(Atmospheric Conditions) (Air Intake Tube)

where $Z$ = elevation, and since $Z_1 = Z_2$, there is no difference in elevation

$P_1$ = atmospheric pressure, inches of water

$P_2$ = pressure in air intake tube, inches of water

$V_1$ = velocity of air in atmosphere = 0

$V_2$ = velocity of air in intake tube, length/time

$\gamma_A$ = specific weight (density of air at ambient conditions), mass/volume

$g$ = force of gravity, length/time/time.

Solving for $V_2$, the air velocity within the air intake tube:

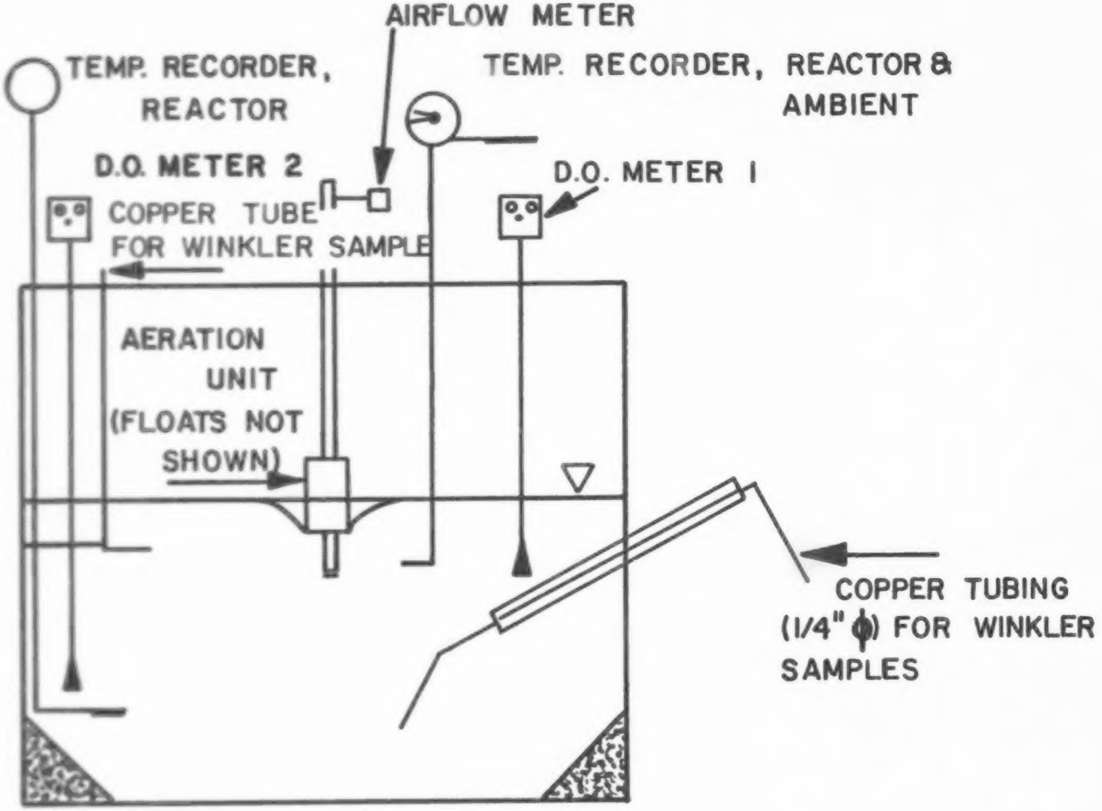

AIRFLOW METER
TEMP. RECORDER, REACTOR
TEMP. RECORDER, REACTOR & AMBIENT
D.O. METER 2
D.O. METER 1
COPPER TUBE FOR WINKLER SAMPLE
AERATION UNIT (FLOATS NOT SHOWN)
COPPER TUBING (1/4" ϕ) FOR WINKLER SAMPLES

Figure 37. Schematic diagram showing the reactor with aeration testing equipment in place.

$$\therefore \quad V_2^{\,2} = \left[\frac{P_1 - P_2}{\gamma_A}\right] 2g \qquad (16)$$

where $P_1 - P_2 = \Delta P$ = actual manometer reading, inches of water.

Correcting the velocity for a square edged orifice, a factor of 0.61, and converting to give units of cm/sec (2g is integrated into the conversion factor), the resulting equation gives the air velocity within the air intake tube.

$$V = (0.61) \left[\frac{(4.93 \times 10^6)\ \Delta P}{\gamma_A}\right]^{\frac{1}{2}} \qquad (17)$$

Airflow into the air intake tube may be calculated from the continuity equation for conservation of mass.

$$Q_A = A\ V \qquad (18)$$

where $Q_A$ = airflow rate, volume/time

$A$ = cross-sectional area of the air intake tube, length$^2$

$V$ = air velocity within the air intake tube, length/time.

Combining Equations (17) and (18) yields an equation for determining the airflow into the system.

$$Q_A = (0.61)\ A \left[\frac{(4.93 \times 10^6)\ \Delta P}{\gamma_A}\right]^{\frac{1}{2}} \qquad (19)$$

where $\gamma_A$ can be found in thermodynamic tables for a given temperature and barometric pressure.

The second method for monitoring airflow into the system was to use an airflow rate meter (Datametrics Model 800-2) with output signal given in units of length per unit time or velocity; therefore, to obtain the volume of air entering the system, the reading from the meter is multiplied by the cross-sectional area of the air intake tube.

These procedures were repeated at reactor temperatures of 60°C, 45°C, and again at 23°C. The D.O. was monitored by two meters, an Orbisphere and a YSI model, at 45°C and 23°C instead of the Winkler method used at the higher temperature. The reason for the procedural change was that no significant difference was observed between the Winkler-determined and the Orbisphere meter-determined values of the reaeration coefficient at a temperature of 60°C.

The oxygen transfer coefficient, $K_{LA}$, for each of the three temperatures studied was calculated using the following equation:

$$dc/dt = K_{LA}(C_s^* - C_t) \qquad (20)$$

where $dc/dt$ = change in D.O. concentration with time

$K_{LA}$ = overall oxygen transfer coefficient, $\text{time}^{-1}$

$C_s^*$ = oxygen saturation concentration at the given conditions, mass/volume

$C_t$ = concentration of D.O. at time t, mass/volume.

$K_{LA}$ was determined by a semi-logarithmic plot of D.O. deficit, $C_s^* - C_t$, vs. time, and a subsequent least squares exponential regression program was used to fit a straight line to the experimental data and compute the slope of the line, $K_{LA}$. The values of $C_s^*$ used in the calculations were taken from Standard Methods (77) at the temperatures of 45°C and 23°C and corrected for local barometric pressure (taken from the aneroid barometer located on the reactor). The value of $C_s^*$ at 60°C was determined from tables provided by Orbisphere Corporation, which in turn, were derived by interpolation between solubilities calculated from the absorption coefficients quoted by Morrison and Billet (82). All tabular values (82) of $C_s^*$ were corrected for local barometric pressure using Equation (21).

$$C_s^* = C_s \left[\frac{P_t - P_w}{760 - P_w}\right] \qquad (21)$$

where $C_s^*$ = actual saturation concentration of D.O. at test conditions, mg/ℓ

$C_s$ = saturation concentration of D.O. at a barometric pressure of 760 mm Hg and a given temperature (tabular value), mg/ℓ

$P_t$ = barometric pressure at test conditions, mm Hg

$P_w$ = vapor pressure of water at test temperature, mm Hg.

The oxygenation capacity of the system was calculated from the following equation:

$$N = (K_{LA\ T°C})\,(C_s^*)\,(\forall) \qquad (22)$$

where $N$ = oxygenation capacity, mass oxygen/time

$K_{LA\ T°C}$ = overall oxygen transfer coefficient at a given temperature, $\text{time}^{-1}$

$C_s^*$ = oxygen saturation concentration at a given temperature, mass/volume

$\forall$ = volume of water under aeration, volume.

The oxygen transfer rate (OTR) was then calculated as follows:

$$\text{O.T.R.} = N/P \tag{23}$$

where

O.T.R. = oxygen transfer rate of the aerator, mass/power-time

$N$ = oxygenation capacity, mass oxygen/time

$P$ = power drawn by aerator, power.

The efficiency of oxygen transfer for the DeLaval centri-rator in clean water was calculated using the equation shown below:

$$\text{O.T.E.} = \left[\frac{N}{(Q_A)\ (\gamma_A)\ (F)}\right] 100 \tag{24}$$

where

O.T.E. = oxygen transfer efficiency, %

$N$ = oxygenation capacity, mass oxygen/time

$Q_A$ = airflow into the system, volume/time

$\gamma_A$ = density of ambient air at given conditions, mass/volume

$F$ = fraction of atmospheric air that is composed of oxygen $\doteq$ 0.2095.

The thermal characteristics of the system were studied in conjunction with the aerator performance tests. Heat losses from the reactor were recorded under aerating and non-aerating conditions as the reactor temperature decreased from 65°C to 20°C by the Partlow temperature recorders described in Section 5. From these heat loss measurements, two important heat loss coefficients could be calculated, $H_L$ and $H_A$. As discussed in Section 4, the overall heat balance may be expressed as follows:

$$\text{BHP} + H_M = H_E + H_V + H_S + H_A \tag{6}$$

where

BHP = biological heat production, °C/day

$H_M$ = heat input due to mixing, °C/day

$H_E$ = rate of heat loss to the effluent sludge, °C/day

$H_V$ = rate of heat loss as latent heat in water vapor, °C/day

$H_S$ = rate of heat loss as sensible heat in the effluent gas, °C/day

$H_A$ = rate of heat loss through the surface area of the reactor, °C/day.

A graphic illustration of the components of the heat balance was presented in Figure 6. Many of the terms presented in Equation (6) are difficult to measure or calculate; however, a system specific heat loss coefficient ($H_L$) and energy balance were developed from the reactor cooling data in hot tap water under aerating conditions. Four components of the heat balance were measured simultaneously, $H_M$, $H_V$, $H_S$, and $H_A$, while the hot tap water was cooling under aeration. The following equation expresses all four terms as one coefficient.

$$H_L = H_V + H_S + H_A - H_M \tag{7}$$

From previous studies (6,7,19, 76), it has been shown that a relationship exists between the heat loss from the reactor and the temperature differential between ambient and reactor temperature. Heat loss was also shown to be dependent upon the density and specific heat capacity of the liquid under aeration. Hence, the following equation:

$$H_L = a(T_R - T_A)^b D_L \; C_{PL} \tag{7a}$$

where
$H_L$ = overall heat loss coefficient, °C/day
$T_R$ = reactor temperature, °C
$T_A$ = ambient temperature, °C
$D_L$ = liquid density, gm/ℓ
$C_{PL}$ = specific heat capacity of liquid, cal/gm-°C
a = intercept coefficient
b = slope coefficient.

A log-log plot of the heat loss from the reactor versus the reactor temperature minus the ambient temperature and a subsequent least squares regression analysis were used to determine (a) and (b) of the above equation.

A term also of interest was $H_A$, the heat loss through surface area of the reactor. This term was calculated in a manner identical to the determination of $H_L$, with the exception that heat loss was monitored under non-aerating conditions. (Note that under non-aeration, the only heat loss from the reactor would be through the surface of the reactor.) The equation for $H_A$ follows:

$$H_A = a(T_R - T_A)^b D_L \; C_{PL} \tag{25}$$

where
$H_A$ = rate of heat loss through the surface area of the reactor, °C/day.

From the determination of $H_A$, the relative importance of reactor insulation could be ascertained.

One other important term was determined from the thermal loss studies using tap water. That term expressed the influence of aeration on heat loss from an aerating liquid. From equation (7), $H_V$, $H_S$, and $H_M$ can be combined into one term, $H_{LA}$, the heat loss due to aeration when $H_L$ and $H_A$ are known. The equation for $H_{LA}$ is shown below.

$$H_{LA} = H_L - H_A \tag{26}$$

It is pertinent to note here that the terms of the heat balance were measured as °C/day; however, one kilocalorie is required to raise the temperature of one liter of water one degree centigrade. The terms °C/day and Kcal/ℓ-day may then be used interchangeably.

## BATCH BIODEGRADABILITY STUDIES

The importance of substrate biodegradability in the autothermal process has been shown by many investigators to be very important (2,6,7,11,19,21,76). Another important consideration in the design of an autoheated digestion process that utilizes sewage sludge is the variability of sludge as related to the time of year and to the mode of STP operation. A series of 14 long-term ($\geq$35 days of continuous operation) batch biodegradability studies were conducted, using the three liter reactors described in Section 5, to measure the maximum biodegradable fraction of the sewage sludge and to monitor the variability of the sludge.

Each reactor was filled with thickened primary and secondary sludge from the Binghamton-Johnson City STP and aerated continuously for at least 35 days in most cases. Temperature, pH, D.O., airflow, and evaporation were monitored daily. Water was added daily to compensate for evaporation losses. At Day 0, 1, 2, 7, 14, 21, 28, and 35, samples were extracted from the reactors and analyzed for the following: Total Solids (TS), Total Volatile Solids (TVS), Total Chemical Oxygen Demand (TCOD), Total Kjeldahl Nitrogen (TKN), Ammonia Nitrogen ($NH_4$-N), Capillary Suction Time (CST), and Zone Settling Rate (ZSR). Only the reactors in the first study were seeded with a small amount of effluent from the full scale thermophilic digester.

The results of the batch studies were used to determine the biodegradable fraction of TS, TVS, TCOD, and organic nitrogen and to monitor the physical changes in the sludge with time (CST and ZSR). Two graphical methods were used in the analyses of the results, the first being a Cartesan plot of the parameter (TS, TVS, TCOD, or organic nitrogen) concentration versus time, as shown in Figure 38. As time approaches forty days, the biodegradable fraction of the sludge approaches zero for all practical purposes. Therefore, the asymtote of the X-axis of time intersects a value on the Y-axis or parameter concentration that is equal to the non-biodegradable fraction or refractory fraction (R) of the parameter plotted. The second method used a technique whereby the refractory fraction (R) could be determined by plotting $S_t/S_o$ versus $(S_o \cdot SRT)^{-1}$ and the following holds:

$S_o$ = substrate concentration at time = zero, mass/volume
$S_t$ = substrate concentration at time = t, mass/volume
SRT = solids retention time, day.

This technique is based on the assumption that as the SRT approaches infinity, the biodegradable fraction of the influent approaches zero; therefore, the

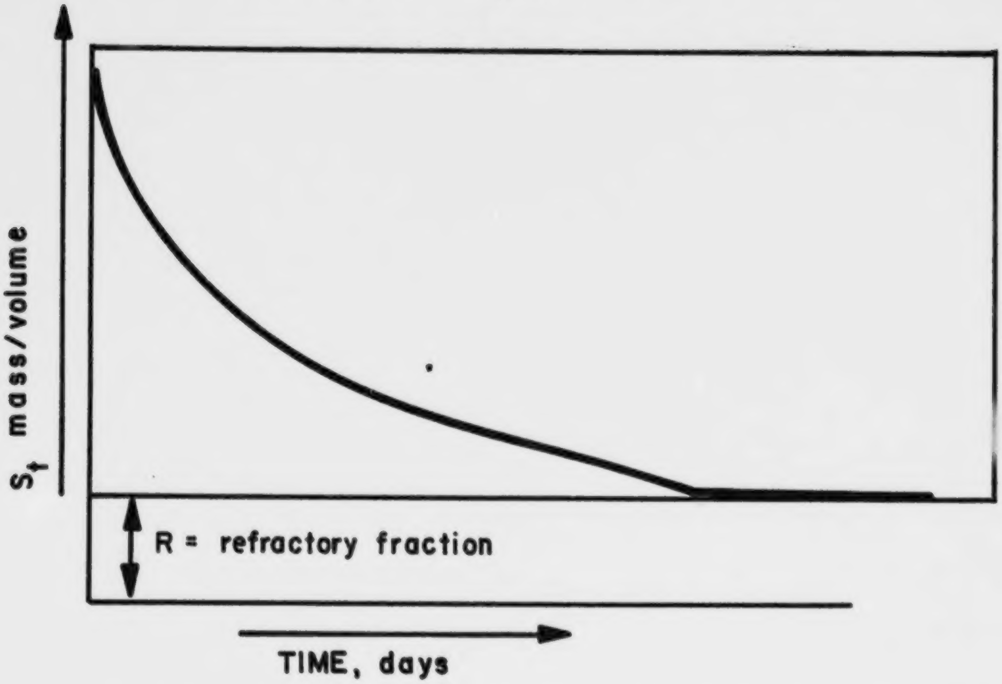


Figure 38. Graphical determination of the refractory fraction of a substrate by Cartesan plot.

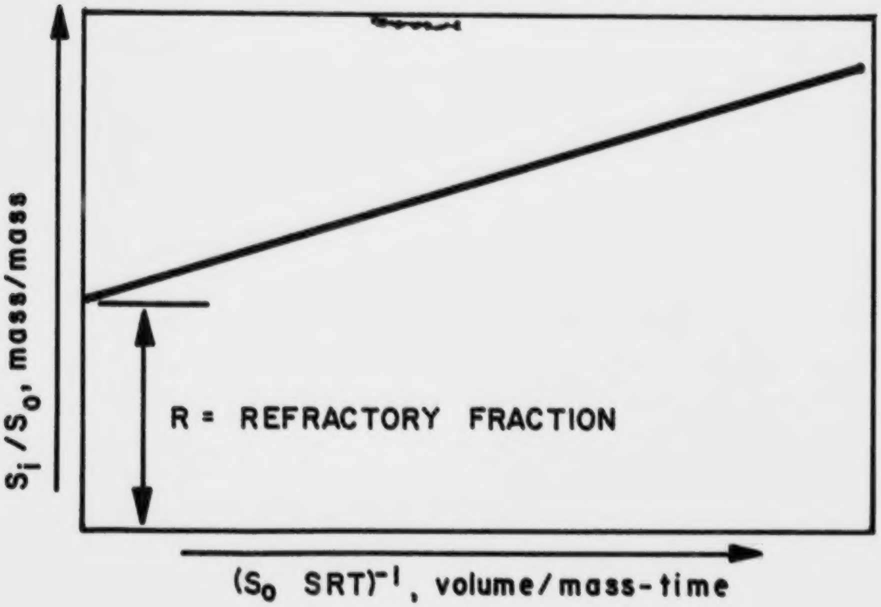


Figure 39. Graphical determination of the refractory fraction of a substrate by reciprocal plot.

## SYSTEMS OPERATION AND DATA COLLECTION

The full scale system and the bench scale continuous and semi-continuous systems were operated under conditions which would result in significant biodegradation of the sludge, but would reflect the advantages of high temperature stabilization. The loading rate, in terms of TS and TCOD as controlled by the hydraulic retention time (HRT), was the primary variable. All conditions tested were at relatively high loading rates with the majority of operation at an HRT of five days or less. Several of the conditions studied were maintained for long periods of time, i.e., greater than or equal to 60 days, in order to provide time to evaluate practical operation and maintenance problems as well as to measure the impact of sludge composition variations on the process. In the full scale system, an optimum loading was sought, i.e., a loading which maintained a high temperature and organic treatment efficiency with the bench scale systems used to document the loadings studied at full scale.

### Full Scale

Analytical data for the full scale study were collected from May, 1977 to December, 1978. Throughout this period, reactor, ambient and feed temperatures, hydraulic flow rates (feed rates), and aerator power consumption were monitored daily. Dissolved oxygen and airflow were measured on a regular basis during steady state operation.

At the beginning of each retention period (HRT), the feed tank was filled with sludge and thoroughly mixed and sampled (as described below) to define influent sludge composition. The feed pump (timer activated corresponding to the HRT under study) was activated and the HRT was begun. At the end of the HRT, the reactor and effluent tank were sampled (as described below) to define the system's performance. Note that sampling was performed during steady state conditions as defined below.

Steady state conditions were examined after the system was operated for a minimum of three to five consecutive hydraulic retention periods which served as an acclimation period. A steady state condition was defined when the organic loading rates, organic removal rates, and reactor temperature were constant within 15% during a minimum of three consecutive HRTs. The results of the representative steady state conditions were numerically averaged to give a definition of the loading condition under investigation. Thirty hydraulic retention periods were tested in an effort to define five loading conditions.

Samples from the feed, reactor, and effluent tank were taken with a core-like sampler developed for use on a similar system at Cornell University. Briefly, the sampler consisted of a 4.6 meter length of 5.0 centimeter diameter PVC pipe with the end sealable by a rubber stopper which in turn was pulled into place with a spring-loaded handle at the opposite end of the pipe. Each time sampling of the three tanks was performed, four core samples of the entire tank height were obtained from the three tanks at different locations within each tank. Each group of core samples was composited (separately, of course), and subsequently, a 1.5 liter subsample was extracted from each composite sample. The samples were then transported to the waste management laboratory at ASTARC or to the AWML for analysis within two hours of collection in most cases. All samples were analyzed for the parameters listed in Table 10 and at the frequency noted.

TABLE 10. TYPES OF ANALYSES AND FREQUENCY OF TESTING

| | | SLUDGE | |
|---|---|---|---|
| Parameter | Influent | Reactor | Effluent |
| Volume fed | Daily | N/A | Steady state |
| Rate of feed | Daily | N/A | N/A |
| Aerator conditions | N/A | Steady state | Steady state |
| Temperature | Continuous | Continuous | Steady state |
| TS | Steady state | Steady state | Steady state |
| VS | Steady state | Steady state | Steady state |
| Capillary suction time | Steady state | Steady state | Steady state |
| Sedimentation | Steady state | Steady state | Steady state |
| COD | Steady state | Steady state | Steady state |
| Sludge sedimentation in the unit | N/A | Steady state | N/A |
| Gas composition ($N_2$, $CO_2$, $O_2$) & flow rate | N/A | Steady state | N/A |
| pH | Daily | Daily | Daily |
| Dissolved oxygen | N/A | Daily | N/A |
| $NH_3$ | Steady state | Steady state | Steady state |
| TKN | Steady state | Steady state | Steady state |
| Heavy metals | Occasionally | Occasionally | Occasionally |

Oxygen Transfer Analysis--
Previous studies (2,11,14,16,17,18,19) have shown that the single most important parameter in autothermal process with air aeration is oxygen transfer efficiency. An aerator must be able to provide enough oxygen to the sludge to meet the high oxygen demand of the aerobic thermophilic mixed liquor, yet have airflow low enough to prevent excessive heat stripping from the reactor (see Section 4). Several kinds of analyses can be used to determine the performance of a given aeration system. Three of these techniques were used in this study; therefore, the total amount of oxygen transferred from the atmosphere to the mixed liquor of the reactor was measured in three separate ways.

Stoichiometric considerations--From stoichometric considerations it can be shown that for every gram of TCOD microbially oxidized, one gram of oxygen is required. Since TCOD removal rates, kg/$m^3$-day, are measured at the various conditions under study, the oxygen transfer efficiency of the aerator(s) at a given condition can be calculated.

$$\text{O.T.E.} = \frac{(\text{TCOD RR})\ (\text{V})}{(Q_A)\ ([O_2]_A)\ (\gamma_A)} \times 100 \qquad (27)$$

where O.T.E. = oxygen transfer efficiency, %

TCOD RR = total chemical oxygen demand removal rate, mass/volume-time

V = volume of reactor, volume

$Q_A$ = airflow rate into reactor, volume/time

$[O_2]_A$ = concentration (atmospheric) of $O_2$ at STP which is 0.2095

$\gamma_A$ = specific weight of air at the given ambient conditions, mass/volume.

Fyrite technique--The second method used Fyrite fluid to measure changes in carbon dioxide ($CO_2$) concentrations between the ambient air and the reactor exhaust gas during steady state operation. The assumption was that any changes in $CO_2$ concentrations between the atmosphere and exhaust gas were the result of microbial respiration. Since the quantity of $CO_2$ released during the microbial oxidation of organic compounds is equal to the amount of oxygen ($O_2$) consumed, the oxygen transfer efficiency of the aerator may be calculated.

$$\text{O.T.E.} = \frac{[CO_2]_E - [CO_2]_A}{[O_2]_A} \quad \frac{D_s}{D_d} \times 100 \qquad (28)$$

where O.T.E. = oxygen transfer efficiency, %

$[CO_2]_E$ = exhaust gas concentration of $CO_2$

$[CO_2]_A$ = atmospheric concentration of $CO_2$, assumed negligible

$[O_2]_A$ = atmospheric concentration of $O_2$ at STP, which is 0.2095

$D_s$ = density of saturated air at reactor temperature and barometric pressure, mass/volume

$D_d$ = density of dry air at reactor temperature and barometric pressure, mass/volume.

Gas chromatography--The third and final method was the utilization of gas chromatography to measure oxygen ($O_2$) concentrations in the intake and exhaust gases of the reactor. Any changes between the two $O_2$ concentrations were assumed to be the result of microbial utilization of $O_2$. The gas chromatograph measured $O_2$ concentrations on a dry basis so there was no correction for saturation of air by water as in the Fyrite technique.

$$\text{O.T.E.} = O_{2\,E} / O_{2\,i} \times 100 \qquad (29)$$

where O.T.E. = oxygen transfer efficiency, %

$O_{2\,E}$ = exhaust gas concentration of oxygen

$O_{2\,i}$ = intake gas concentration of oxygen.

### Bench Scale

Analytical data for the bench scale systems were collected during the period from January, 1977 to August, 1978. Dissolved oxygen, pH, airflow, hydraulic flow rates (feed rates), evaporation, and temperatures were monitored daily throughout the duration of reactor operation.

Thickened primary and secondary sewage sludge was collected from the Binghamton-Johnson City STP thickener pumps in a 200 liter container and transported to ASTARC or the AWML to serve as feed for the bench scale system. The sludge was strained through a coarse, 1 cm in diameter, sieve to remove large objects, e.g., cigarette butts, condoms, etc., which might interfere with the feed pump. The sludge was then stored in a plastic-lined and calibrated feed barrel at 0°C for the continuous unit or in a plastic bucket at 0°C for the semi-continuous unit. Each batch of fresh sludge was analyzed after the straining process and regularly throughout the period of its use.

Continuously-fed unit--

The feed rate for the continuous unit was controlled by a timed electric chemical pump (activation interval corresponding to the HRT under study). Since wasting from the reactor was accomplished by gravity overflow, the amount of sludge pumped in (through the bottom of the reactor; see Figure 35), as measured by the displacement in the feed barrel, could be compared to the amount wasted in the effluent collection barrel. Any discrepancies between the measurements were due to evaporation and were compensated for accordingly. This system allowed for the collection of an entire HRT's effluent and approximated the full scale operation. The continuous feed system also prevented shock-loading which would have occurred with the hand feeding of a 3-day HRT reactor, i.e., once per day replacement of one-third of the reactor contents. This unit was used to study an approximate 3-day HRT.

Semi-continuously fed unit--

This unit was hand-fed once or twice per day by research personnel. The feeding procedure consisted of the removal of a given volume of the reactor (after compensation for evaporation) corresponding to the HRT under study and subsequent addition of the same volume of feed sludge. This unit was used to investigate HRTs ranging from 5 to 10 days.

Steady state conditions in both systems were examined as set forth above for the full scale system with a steady state condition defined here just as in the full scale. Twenty-one hydraulic retention periods were tested in an effort to define four loading conditions.

Each reactor and effluent-holding barrel was sampled at the end of a respective HRT. The reactors were grab sampled, and the effluent-holding barrels were thoroughly mixed by hand and subsampled (approximately one to two liters). The samples were usually analyzed immediately after collection. All samples were examined for the parameters listed in Table 10 and at the frequency noted.

## ANALYTICAL METHODS AND DATA CALCULATIONS

Dissolved oxygen measurements were made with a specially prepared dissolved oxygen meter. This instrument was adapted for use in high solids, organic slurries at high temperatures by Orbisphere Laboratories of Geneva, Switzerland, especially for Cornell University. The D.O. meter was calibrated via the azide-modification of the Winkler D.O. analysis (77). Two pH meters, a Fischer scientific portable for the full scale system and an Orion table model for the bench systems, were used for measuring pH immediately upon collection of the sample. A Gow-Mac brand (Model 550) thermal conductivity gas chromatograph utilizing a molecular sieve separation column was used to determine oxygen concentration in the reactor stack gases. A Fyrite fluid instrument was used for determining carbon dioxide concentrations in the stack gases.

Total solids, total volatile solids, and total chemical oxygen demand were determined in accordance with procedures outlined in Standard Methods (77) for the preparation and analysis of semi-solid samples. Total Kjeldahl nitrogen and ammonia-nitrogen were determined by the macro-Kjeldahl method (77) at ASTARC and by the micro-Kjeldahl method at the AWML (83). The zone settling velocity of the sludge was determined by measuring the downward movement of the reactor sludge-supernatant interface with time in a one liter graduated cylinder (standard) equipped with a 10 revolution per hour mixer. The CST (Capillary Suction Time) measurements were made using equipment as described by Baskerville and Gale (84).

Calculations were made, utilizing the analytical data gathered, to determine the performance of the systems in this study. The equations for such calculations are shown below:

$$\mathrm{HRT} = \Theta_H = \forall/Q \qquad (30)$$

where HRT, $\Theta_H$ = hydraulic retention time, time

$\forall$ = volume of the system reactor, volume

$Q$ = flow rate to the system, volume/time.

$$\mathrm{L.R.} = S_o/\Theta_H \qquad (31)$$

where L.R. = loading rate of given waste characteristic, mass/volume-time

$S_o$ = influent waste characteristic, mass/volume.

$$R.R. = S_o - S_e/\theta_H \qquad (32)$$

where $S_e$ = effluent waste characteristic, mass/volume.

$$R.E. = \frac{S_o - S_e}{S_o} \times 100 \qquad (33)$$

where R.E. = removal efficiency (or treatment efficiency) of a given parameter, %.

## PATHOGEN ISOLATION AND DETECTION

As discussed in a previous section, one of the many advantages of thermophilic sludge digestion is the significant reduction in the numbers of pathogenic microorganisms incurred by the process. In order to qualify as well as quantify the reduction of pathogens in the autoheated aerobic digestion process, an extensive literature review and communications with known authorities in the field of study were used to select organisms for study (see theoretical section). The fate of three groups of pathogens, bacteria, viruses, and parasites in the full scale thermophilic aerobic digester and full scale mesophilic anaerobic digesters was investigated during this study. Four laboratories were contracted to perform the analyses for the various pathogens. These agencies are listed below along with the kind(s) of organism(s) they were to identify and/or quantify.

| Contracted Agency | Pathogenic Organism |
|---|---|
| (1) U.S. Environmental Protection Agency<br>Environmental Monitoring & Support Lab.<br>Cincinnati, Ohio | bacteria, viruses |
| (2) Dr. E.S. Kaneshiro, Dep't of Bio. Sciences<br>University of Cincinnati, Cincinnati, Ohio | parasites |
| (3) Dr. M. Dale Little<br>School of Public Health & Tropical Medicine<br>Tulane University, New Orleans, Louisiana | parasites |
| (4) Dr. Sang Shin, Diagnostic Laboratory<br>N.Y.S. Veterinary College<br>Cornell University, Ithaca, New York | bacteria |

During most of the steady state conditions studied with the full scale thermophilic digester, samples were collected from the thickener pumps (raw), the anaerobic effluent, and the aerobic effluent and shipped to the contract laboratories. The following section of this report describes the sample collection and shipment, and the various analytical techniques used to quantify and/or qualify the pathogenic microorganisms under study.

## Sample Collection and Shipment

Pathogen sample collection and shipment were always conducted on a Monday to ensure that the contract laboratories were able to analyze the samples, especially the bacteria samples, the same week of collection. Typically, sludge samples, approximate volume of 10 liters, were collected from the three sources: raw sludge from the thickener pumps, aerobic sludge from the experimental thermophilic reactor, and anaerobic sludge from the existing STP digester (specifically, samples were obtained from the draw-off valve used to fill the tanker-spreaders used in land application of the sludge from the anaerobic digester). Some of the samples from the aerobic digester (samples were taken from the bottom drain valve and/or taken from the top of the reactor through the sample ports) were collected under batch-mode conditions, i.e., the feed pump was disconnected for 24 hours prior to sample collection to mimic a system that was being operated to ensure optimum pathogen kills.

After the collection of the gross samples, 250 ml subsamples were extracted and placed in sterilized (autoclaved at 151°C and 15 p.s.i. for 15 minutes) nalgene bottles and sealed. Typically, each of five sets of these subsamples, consisting of one set for virus analysis, two sets for parasite analysis, and two sets for bacterial analysis, were placed in insulated shipping containers (Trans. Temp. System 509) with cold packs (note that the virus samples were frozen with dry ice at -70°C in the shipping containers) and sealed. The shipping containers were appropriately labeled and addressed, then dispatched via air freight (Airborne Freight Corp.) to the various laboratories.

## Analytical Methods

All of the analyses for virus in the sludge samples were performed by the Environmental Monitoring and Support Laboratory of the U.S. EPA. Methods similar to those developed by Berg (85) were employed; however, the exact methodology is still under development and refinement, and has not been published to date (86). No description can be rendered in this report (86).

The analyses for parasites were performed by two separate laboratories using two different isolation and detection schemes. Tulane University, School of Public Health and Tropical Medicine, employed techniques developed for a separate study under EPA contract (87). These procedures are presented in Appendix A-1. The University of Cincinnati analyzed for *Ascaris* sp. only, using a technique also developed under EPA contract (89). A summary of the methodology is presented in Appendix A-2.

The U.S. EPA Environmental Monitoring and Support Laboratory performed analyses for fecal coliform, total coliform, and fecal streptococci in accordance with *Standard Methods* (77) and *Microbiological Methods for Monitoring the Environment* (89). Analysis was also made for enteric pathogens, specifically *Salmonella* sp. by use of the modified Kenner-Clark procedure for enumerating *Salmonella* sp. (90). This procedure was also used to enumerate *Pseudomonas aeruginosa*, an opportunistic pathogen. Sludge samples were screened by the Cornell University Diagnostic Laboratory, School of Veterinary Medicine, for common enteric pathogens, e.g., *Salmonella*, *Shigella*, *Arizona*, and *Edwardsiella* in accordance with procedures developed by Edwards and

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Ewing (91). The primary enrichment broths were Duciltol Selenite Broth (Difco) and GN Broth (Difco) with brilliant green agar and xylose lysine desoxycholate agar used as selective media, respectively. Conformation of pathogens was made via colonial morphology, biochemical tests, and seriological agglutination. (See Appendix A-3.)

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# Section 7

# RESULTS

This study was conducted to illustrate the possibilities of achieving high temperatures in liquid sewage sludge treatment for enhancement of aerobic stabilization at a low cost. By using a full scale system, it was intended to demonstrate the practicality of an autoheated process with air aeration under real world conditions and to provide supporting information that would illustrate the limits of this unique approach by defining the controlling fundamentals. This project followed a three-year study of the process as it applied to animal wastes at Cornell University (6,7,19) in which autoheating with air aeration was successfully demonstrated. The study reported here covered a thirty-month period beginning in October, 1976 and running through April, 1979. In general, the major results are summarized in this section, and complete operating data are given in the Appendices. Table 11 is an outline of the major activities of the full scale and bench scale experiments.

## DEFINITION OF THERMAL CHARACTERISTICS; OXYGEN TRANSFER (AERATOR) - CLEAN WATER

In order to achieve autoheating of a liquid organic slurry, it was known that the reactor would have to be well insulated and the aerator would have to be capable of achieving exceptionally high oxygen transfer efficiency. The standard approach of defining the aerator characteristics in tap water and then modifying them for application to the organic slurry was used here.

The performance of the Delaval centri-rator in clean tap water and at elevated temperatures was examined to provide a basis for evaluating aerator performance. This procedure of evaluating aerator performance under nonsteady state clean water conditions, and subsequently, applying these determinations, with appropriate conversion factors, to field applications is the common method of aerator design and application. However, as was found by Cummings and Jewell (6, 7), the centri-rator was found to perform poorly in tap water. The results of tap water tests are presented in Table 12. The maximum observed oxygen transfer efficiency (O.T.E.) was 3.45%; the oxygen transfer efficiency decreased to around 2% at 60°C. This was as would be expected due to the lower liquid viscosity at high temperatures. (The actual logarithmic deficit plots for the determination of $K_{LA}$ are found in Appendix B.)

The actual performance of the centri-rator under steady state conditions, as will be shown in a subsequent section, greatly surpassed that observed in the clean water tests. This was expected because a similar phenomenon was observed using a dairy waste substrate with the centri-rator (6, 7). Possible reasons for this enhancement of oxygen transfer will be discussed later. It

## TABLE 11. MAJOR PROJECT ACTIVITIES

| Dates | Activity |
|---|---|
| October 1976-January 1977 | -Project initiation |
| February 1977-March 1977 | -Full and bench scale system design; equipment and material procurement; operation of existing full scale AATD on dairy waste for orientation of personnel; bench scale start up, and debugging, high temperature problems. |
| November 1977-May 1977 | -Procurement of equipment, and construction of new full scale digester facility at the Binghamton-Johnson City STP. |
| May 1977-September 1977 | -Full scale start up and debugging; continuous operation of full scale and bench scale systems at thermophilic conditions; biodegradability studies begin. |
| Sept. 1977-Nov. 1977 | -System modification; thermal and oxygen transfer studies. |
| Nov. 1977-Dec. 1977 | -Procurement of winterization materials; freezing problems with full scale unit caused shutdown. |
| Dec. 1977-Feb. 1978 | -Freezing problems continue; winterization completed. |
| February 1978-May 1978 | -Successful continuous operation of full and bench scale systems under cold conditions. |
| June 1978-July 1978 | -Feed pump problems (full scale); preliminary testing of second aerator, Midland-Frings unit. |
| July 1978-August 1978 | -Completion of all bench scale work; Midland-Frings aerator operation with full scale at high loading rates. |
| Sept. 1978-Dec. 1978 | -Full scale system operation at long retention for final data development and pathogen studies. |
| December 1978-April 1979 | -Data analyses; report development. |

## TABLE 12. RESULTS OF NONSTEADY STATE CLEAN WATER TESTING OF CENTRI-RATOR AERATION SYSTEM[+]

| | Reactor Temperature (°C) | | | |
|---|---|---|---|---|
| | 23.0<br>(D.O. meter) | 45.0<br>(D.O. meter) | 60.0<br>(Winkler) | 60.0<br>(D.O. meter) |
| $C_s^*$ (mg/ℓ) | 8.34 | 5.94 | 4.63 | 4.63 |
| $K_{LA}$ ($hr^{-1}$) | 4.73 | 6.72 | 7.81 | 7.91 |
| Correlation Coefficient ($r^2$) | 0.998 | 0.992 | 0.985 | 0.996 |
| Airflow (ℓ/sec) | 37.1 | 34.8 | 37.1 | 37.1 |
| Power (kW) | 3.31 | 3.37 | 3.22 | 3.22 |
| Oxygen Transfer Rate (kg/kW-hr)[#] | 0.338 | 0.336 | 0.319 | 0.323 |
| Oxygen Transfer Efficiency (%) | 3.19 | 3.45 | 1.80 | 2.97 |

[+] Results are averages of duplicate runs.

[#] $$\frac{\text{kg } O_2}{\text{hr}} = C_s^*\text{, mg/ℓ} \times K_{LA}\text{, hr}^{-1} \times \text{reactor vol., 28,400 1} \times 10^{-6}\text{ kg/mg.}$$

is apparent that tap water evaluation of the Delaval centri-rator, and possibly other submerged, self-aspirating aerators, in clean tap water provides little or no basis for predicting subsequent performance in organic slurries at elevated temperatures. For this reason, no further tap water analysis of the aerators was conducted, and no testing of the second aerator, the Midland-Frings unit, was completed.

The thermal characteristics of the system were measured by monitoring the heat losses from the full scale reactor in response to changes in the ambient air temperature. From the reactor cooling data, at relative constant ambient temperatures (since any changes in the ambient temperature subsequently affect reactor temperature), heat loss equations for static (nonaerating) and dynamic (aerating) conditions were developed with tap water. Although it was anticipated that the cooling characteristics with the organic slurry might differ from tap water, heat losses with a reacting organic slurry would be complicated by the exothermic biochemical reactions that would occur. Thus, the heat losses defined by the tap water tests can only be used as approximations of the operating system with sewage sludge. Typical temperature decreases in aerating and quiescent reactor conditions are shown in Figure 40.

The heat losses from the system and the corresponding temperature differential between the reactor and ambient air temperature when plotted on logarithmic paper yield a straight line, thereby producing an equation(s) in the form of $y = ax^b$. It is important to note these heat loss equations are system specific and caution should be used when applying the equations to other systems. The resulting heat loss equations for aerating ($H_L$) and nonaerating ($H_A$) conditions are shown below with the actual data tabulated in Tables 13 and 14 and graphically depicted in Figure 41.

$$H_L = 0.0407\ (T_R - T_A)^{1.28}\ D_L C_{PL} \qquad (34)$$

$$H_A = 2.74 \times 10^{-4}\ (T_R - T_A)^{2.28}\ D_L C_{PL} \qquad (35)$$

where $H_L$ = overall heat loss coefficient, °C/day

$H_A$ = rate of heat loss through surface area of reactor, °C/day

$T_R$ = reactor temperature, °C

$T_A$ = ambient temperature, °C

$D_L$ = liquid density, gm/ℓ ($\simeq$1.0 for water and sewage sludge at 5.0% total solids)

$C_{PL}$ = specific heat capacity of liquid, cal/gm -°C ($\simeq$1.0 for water and assumed same for sewage sludge at 5.0% total solids).

Since the overall heat balance for the system can be expressed as follows:

$$BHP = H_L + H_E \qquad (36)$$

where BHP = biological heat production, Kcal/ℓ-day

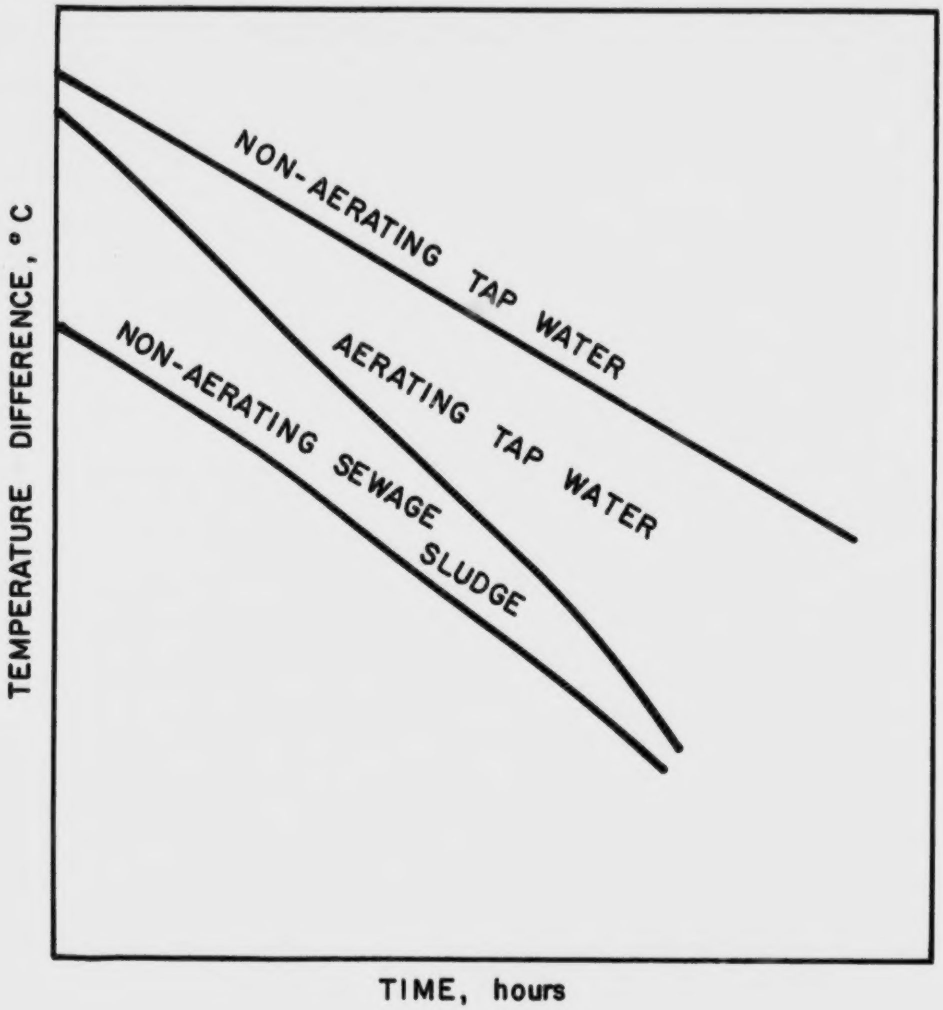

Figure 40. Typical cooling in the full scale aerobic digester.

TABLE 13. DETERMINATION OF THE HEAT LOSS CHARACTERISTICS OF THE REACTOR DURING AERATION OF TAP WATER*

| Pt | Temp Reactor Initial °C | Temp Reactor Final °C | Temp Reactor Average °C | Temp Ambient Average °C | Average $T_R-T_A$ °C | Heat Loss °C/day |
|---|---|---|---|---|---|---|
| 1 | 56.5 | 55.5 | 55.8 | 16.6 | 39.2 | 4.80 |
| 2 | 47.5 | 46.5 | 47.0 | 14.4 | 32.6 | 3.43 |
| 3 | 41.0 | 39.0 | 39.9 | 8.5 | 31.4 | 3.20 |
| 4 | 37.0 | 36.0 | 36.4 | 9.5 | 26.9 | 2.40 |
| 5 | 35.5 | 34.5 | 35.0 | 13.1 | 21.9 | 2.18 |
| 6 | 34.0 | 33.5 | 33.8 | 18.1 | 18.1 | 1.17 |

*Linear regression analysis (power curve) yielded the following:

$$H_L = 0.0407\ (T_R-T_A)^{1.28},\ r^2 = 0.9586.$$

TABLE 14. DETERMINATION OF THE HEAT LOSS CHARACTERISTICS OF THE REACTOR DURING QUIESCENT CONDITIONS*

| Pt | Temp Reactor Initial °C | Temp Reactor Final °C | Temp Reactor Average °C | Temp Ambient Average °C | Average $T_R-T_A$ °C | Heat Loss °C/day |
|---|---|---|---|---|---|---|
| 1 | 55.0 | 54.5 | 54.8 | 16.6 | 38.2 | 1.50 |
| 2 | 46.0 | 45.0 | 45.5 | 13.5 | 32.0 | 1.09 |
| 3 | 45.0 | 44.5 | 44.8 | 13.4 | 31.4 | 1.00 |
| 4 | 31.0 | 30.5 | 30.8 | 13.7 | 29.4 | 0.80 |

*Linear regression analysis (power curve) yielded the following:

$$H_A = 3.74 \times 10^{-4}\ (T_R-T_A)^{2.28},\ r^2 = 0.9625$$

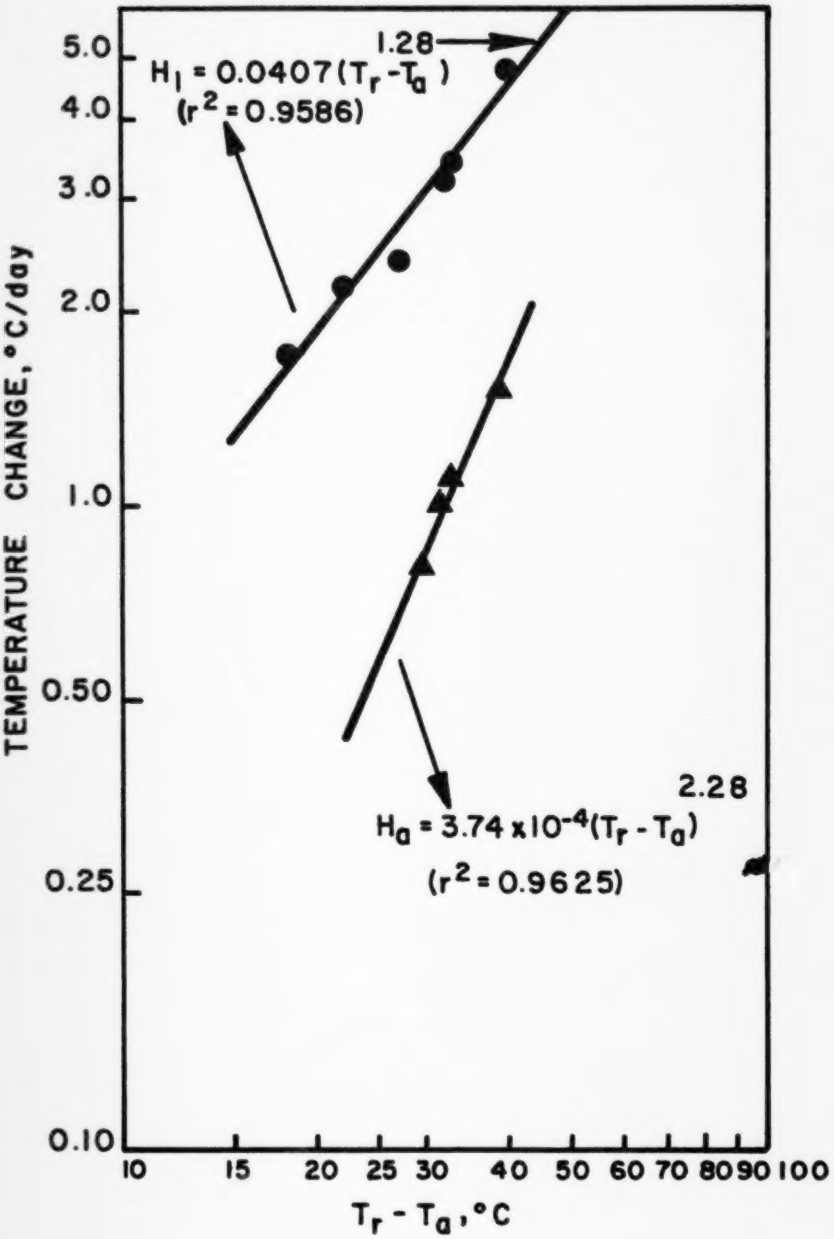

Figure 41. Static and dynamic heat losses as related to the average temperature difference (reactor and ambient) for full scale, non-steady state, tap water tests.

$H_E$ = heat lost to effluent sludge, °C/day

$$H_E = \frac{T_R T_o}{HRT} D_L C_{PL} \qquad (8)$$

$T_R$ = reactor temperature, °C

$T_o$ = feed sludge temperature, °C

HRT = hydraulic retention time, day,

it can be seen $H_L$ is important not only in defining the overall heat loss but also in determining biological heat production under steady state conditions. The evaluation of the major thermodynamic variables, $H_L$, $H_A$, $H_{LA}$, and $H_E$ under actual steady state operation with sewage sludge (autoheated) will be presented later in this section.

Although for calculation purposes it was assumed that the heat loss curve for the reactor filled with sewage sludge would be the same as those determined from tap water, it must be realized that the heat losses with aerated sewage sludge were probably lower than those in tap water due to two factors: under operation with sewage sludge the airflow through the aerator was reduced by 10 to 12 $\ell$/sec (or more than 25%) which would mean a savings of up to 4.0 Kcal/$\ell$-day by decreasing the heat loss due to the latent heat of vaporization ($H_V$) and the sensible heat in the effluent stack gases ($H_S$). The second heat conserving factor was the constant presence of a thick and dense foam layer during steady state operation. Although the influence of this foam layer was not assessed, it may be assumed that the resistance to heat loss of a densely packed foam is at least one order of magnitude higher than that of a liquid surface experiencing constant turnover without a dense foam layer. The implications of these two factors and the importance of other heat balance variables will be discussed in a subsequent section.

## SEWAGE SLUDGE BIODEGRADABILITY ANALYSES: BATCH DECOMPOSITION STUDIES

The achievement of autoheating in aerobic digestion depends upon the total quantity of organics available for biodegradation and the rate at which they can be biologically oxidized. No attempt was made to modify the thickened primary and waste activated sludge developed by the sewage treatment facility (see earlier description of the sludge) prior to its use in the decomposition studies. In order to define the variability of the sludge, laboratory batch decomposition studies were conducted throughout the testing period.

The results of the batch biodegradability studies were used to determine the following:

(1) Physical changes in dewatering characteristics, i.e., capillary suction time and zone settling rate.

(2) Identify patterns of destruction of stated parameters.

(3) Quantify the biodegradable fractions of the parameters used to characterize sewage sludges, e.g., TS, TVS, TCOD, TKN.

(4) Variability of the sludge with respect to time of the year and sewage treatment facility mode of operation.

Typically observed removal patterns, and physical changes in the batch units, are shown in Figures 42 through 46. A complete set of tables and figures for all of the studies is given in Appendix C. A graphical method with linear regression analysis, as described in section 4, was used for determining the biodegradable fraction of TS, TVS, TCOD, and TKN. This method was chosen over the cartesan method (as described in Section 6) due to its more conservative nature as well as its sound statistical base. The results of the seven biodegradability studies are summarized in Table 15.

Dewaterability, as measured by CST and ZSR, was found to improve over initial values only after long periods, ≥30 days, of aeration. This is demonstrated in Figure 47. These results support previous findings that sludge dewaterability (or settleability (as measured by ZSR)) is proportional to sludge age (92). This concept is demonstrated in Figure 47. pH in the batch units was found to vary between 6.6 and 7.6 after 1 day of aeration as shown in Figure 48.

The rate of biodegradability was observed to fluctuate from study to study as well as decrease with time of operation (see Appendix C). The easily biodegradable material was quickly removed with the more difficult material being removed at a slower rate in the latter stages of operation. The effect of this variable rate of biodegradability on continuous, steady state operation will be discussed in a subsequent section.

During the first four biodegradability studies covering the period August 17, 1977 to March 2, 1978, a period of seven months, the observed removal patterns for TCOD, TS, and TVS remained relatively constant with the biodegradable fractions varying only five percent. Observed changes in the physical nature, i.e., pH, CST, ZSR, of the sludge during aeration over the test periods (35 days) also remained constant. This consistency was observed in spite of fluctuations in the raw sludge from the Binghamton-Johnson City STP, e.g., fluctuations in the sludge concentration (see previous section).

The last three biodegradability studies covering the period June 5, 1978 to August 29, 1978, a period of three months, showed a marked decrease in the biodegradable fractions of TS, TVS, and TCOD, as well as a high degree of variability. This decrease in biodegradability of the sludge and increase variability was a function of the time of year and the operational characteristics of the sewage treatment facility.

The average results of biodegradability studies 1 through 4 were used to describe steady state operational performance, both full and bench scale, in terms of biodegradable parameters for all steady state conditions tested during the period May, 1977 to June, 1978. An average biodegradability was used in this case because of the fact that no significant difference was observed between the results of studies 1 through 4.

Biodegradability studies 5 through 7 were not averaged because of the large difference between the three studies. The results of each study were applied

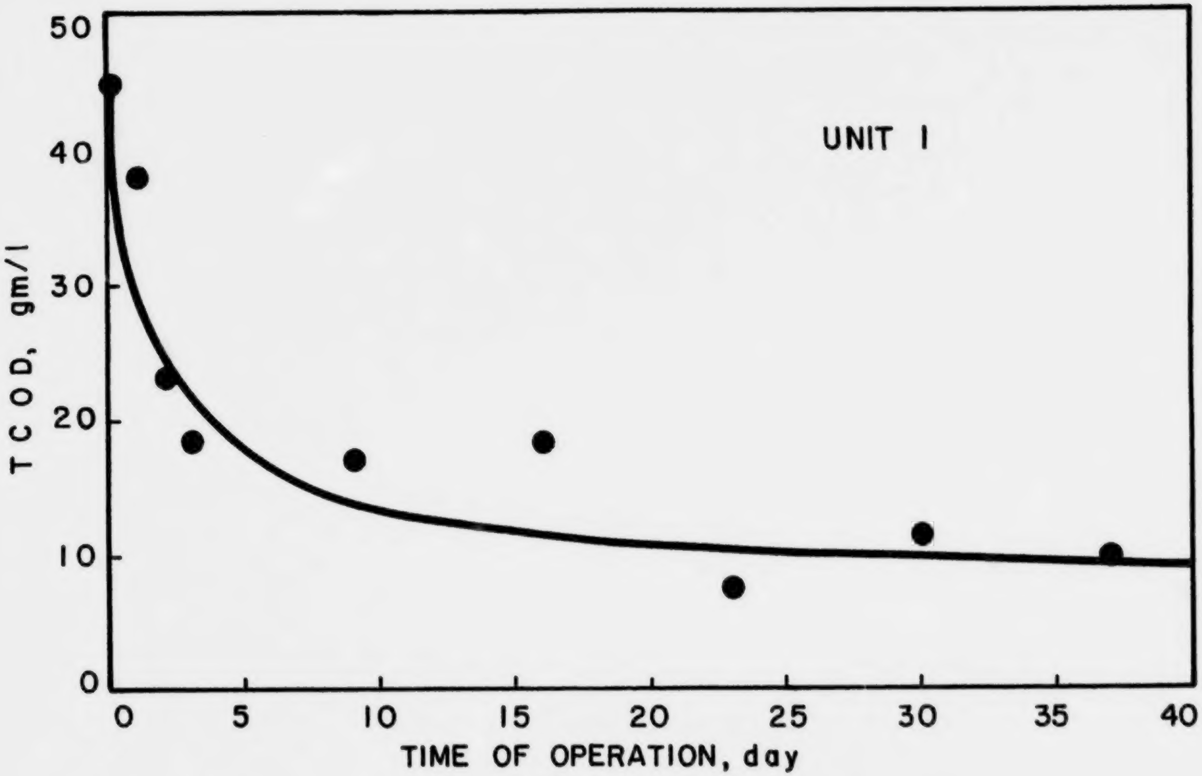

Figure 42. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 4.

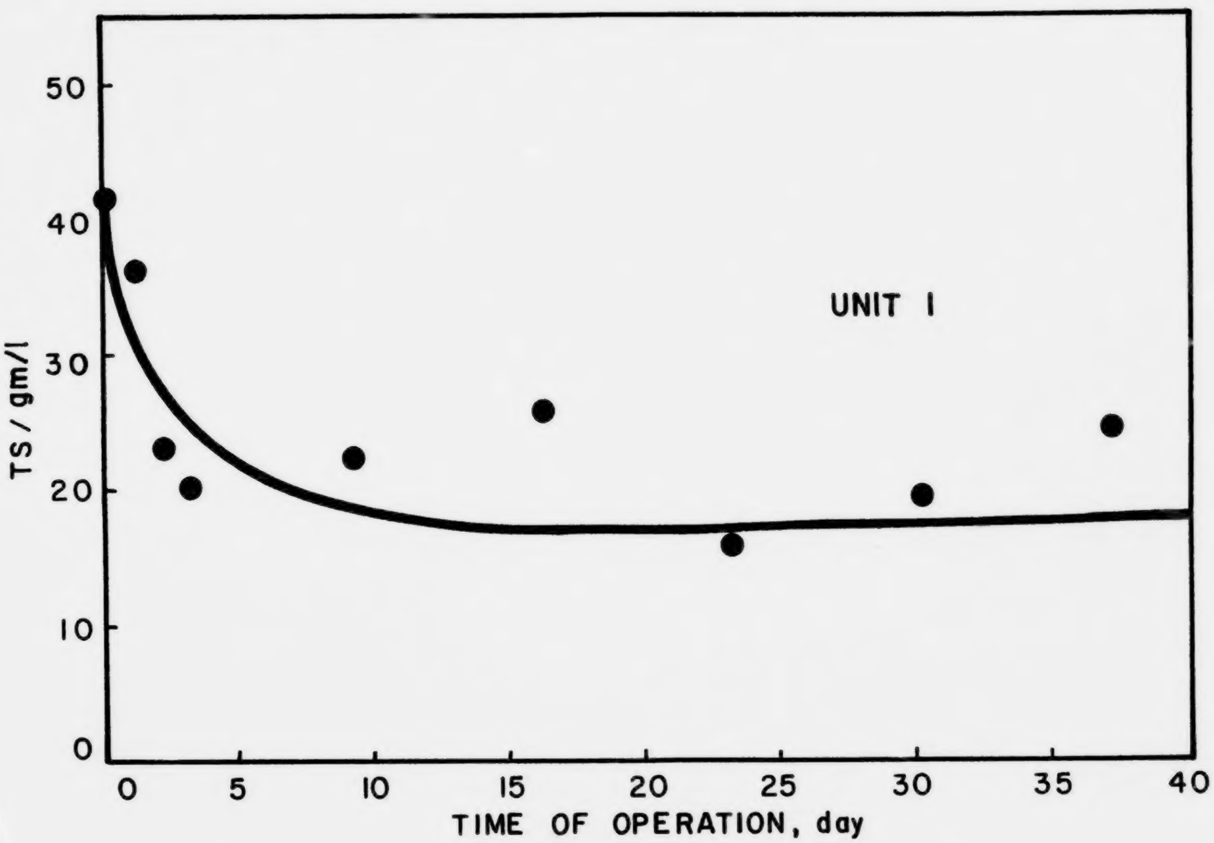

Figure 43. Total solids (TS) removal as related to time of operation, biodegradability study no. 4.

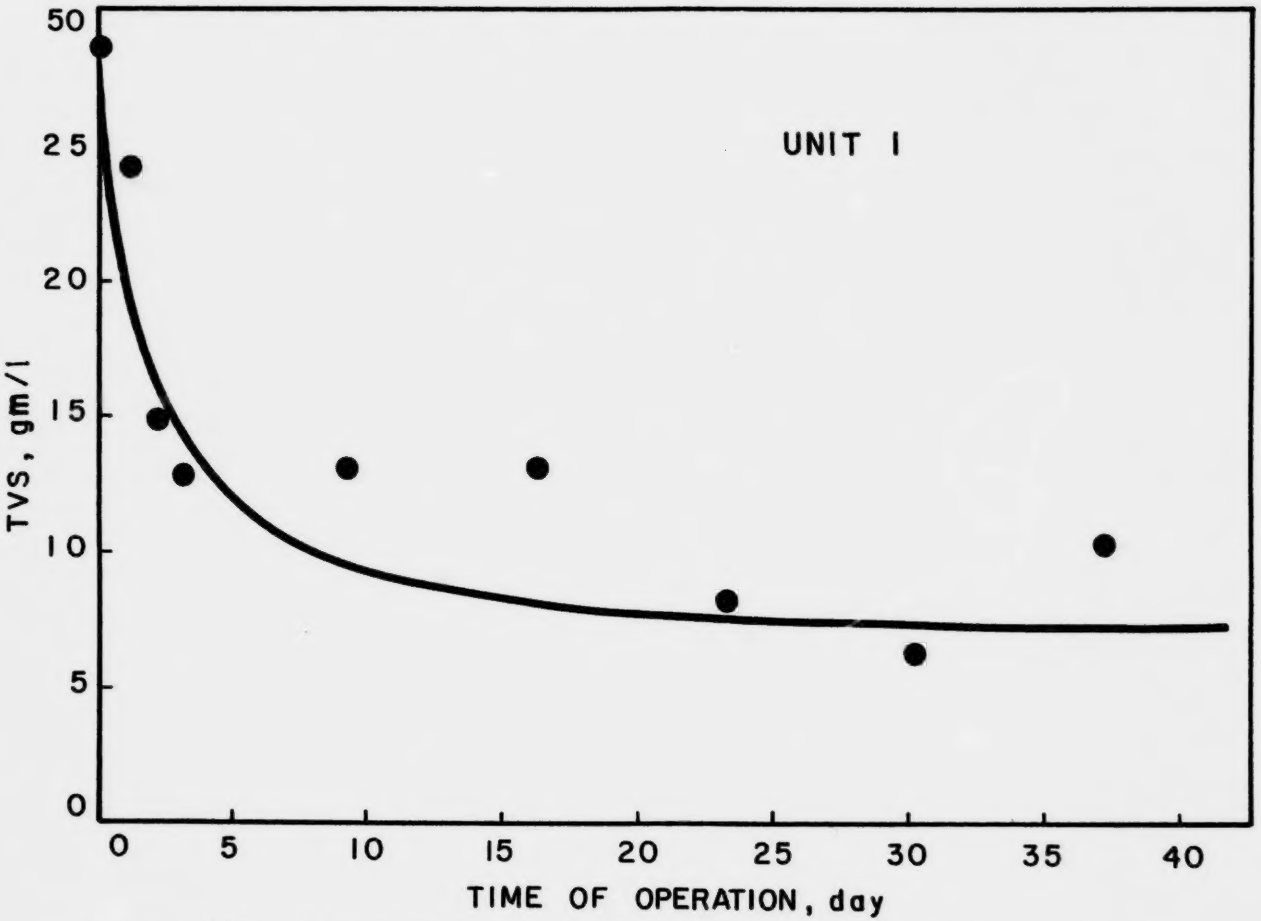

Figure 44. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 4.

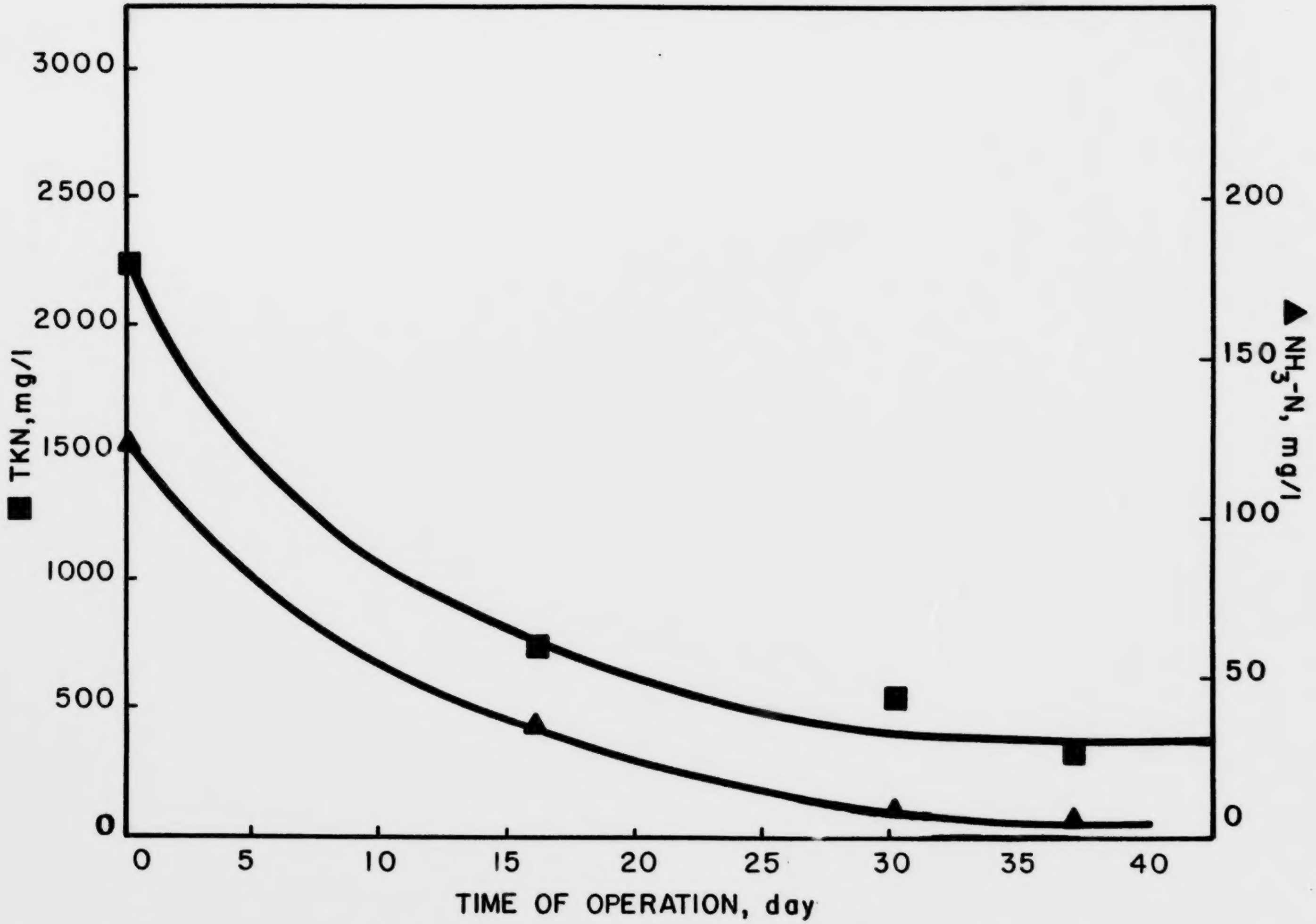

Figure 45. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 4.

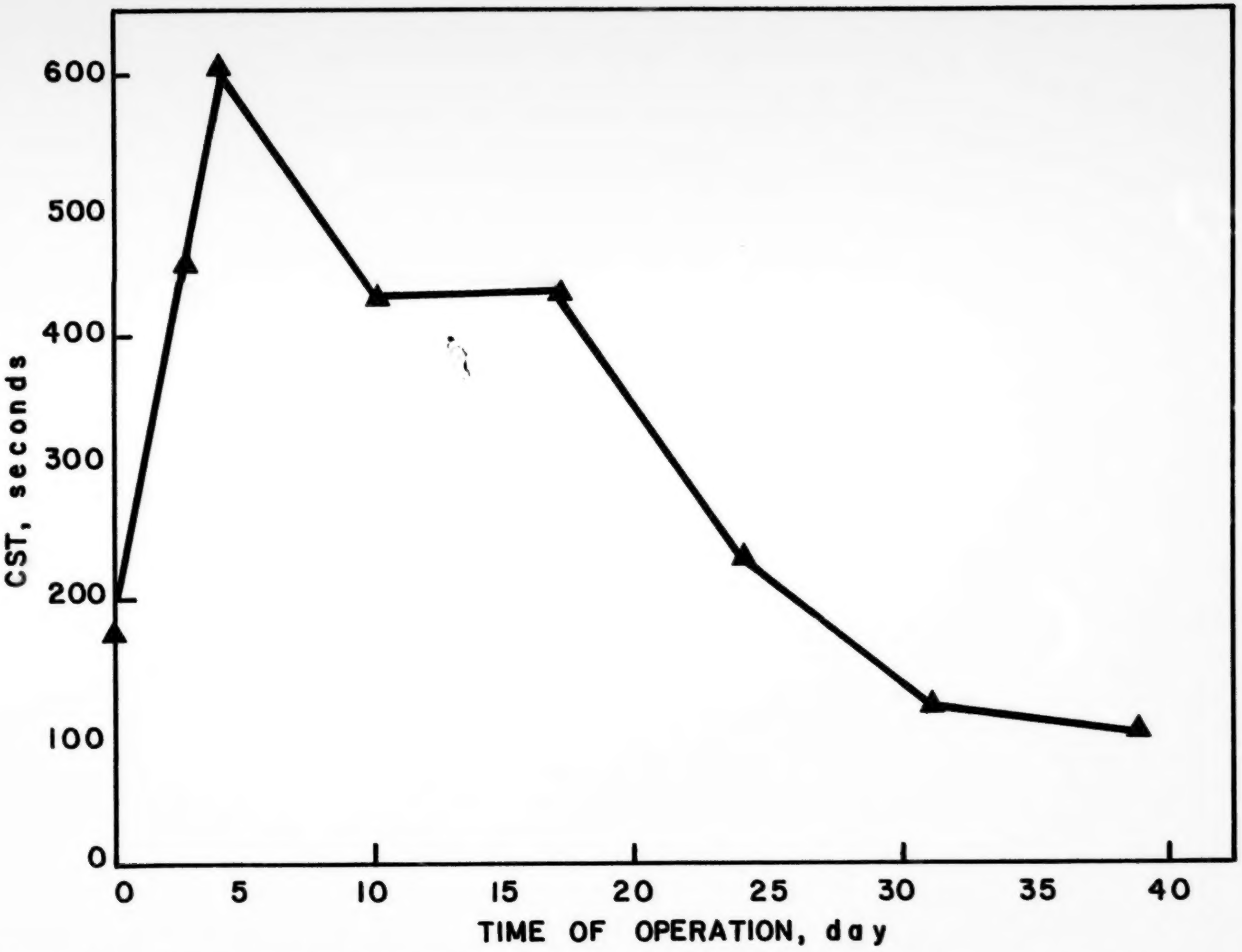

Figure 46. Change in sludge dewaterability (CST) as related to time of operation, biodegradability Study No. 4, Unit No. 1.

TABLE 15. RESULTS OF SEVEN BIODEGRADABILITY STUDIES COVERING THE PERIOD AUGUST 17, 1977 TO AUGUST 29, 1978*

| No. | Dates of Operation | Average Reactor Temp °C | Average D.O. mg/ℓ | Biodegradable Portion† COD % | TS % | TVS % | TKN % | Removal Efficiency $NH_3$-N % |
|---|---|---|---|---|---|---|---|---|
| 1 | 8/17/77-<br>9/19/77 | 52.0 | 4.2 | 75.6 | 60.3 | 67.0 | - - | - - |
| 2 | 9/21/77-<br>10/21/77 | 52.1 | 4.3 | 76.5 | 62.0 | 70.0 | - - | - - |
| 3 | 10/25/77-<br>11/20/77 | 53.5 | 2.0 | 69.0 | 57.0 | 63.5 | - - | - - |
| 4 | 1/24/78-<br>3/02/78 | 51.0 | 3.8 | 75.7 | 58.1 | 71.7 | 85.0 | 96.0 |
| 5 | 6/05/78-<br>7/14/78 | 50.5 | 4.1 | 47.4 | 22.9 | 42.2 | 57.0 | 83.7 |
| 6 | 6/28/78-<br>7/31/78 | 52.6 | 4.6 | 60.5 | 35.0 | 53.4 | 59.4 | 82.7 |
| 7 | 8/02/78-<br>8/29/78 | 50.3 | 3.4 | 64.9 | 29.8 | 53.1 | - - | - - |

*Experiments No. 1-7.

†As determined by the refractory plot method.

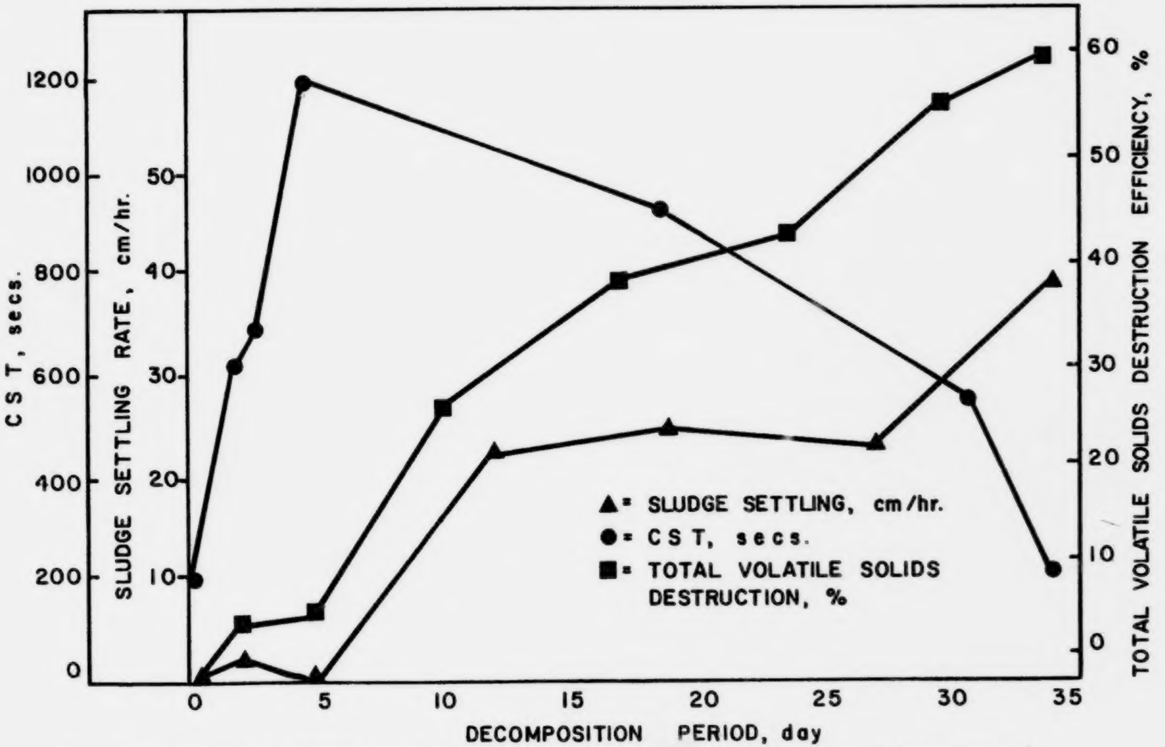

Figure 47. Change in sludge dewaterability, CST and ZSR, as related to time of operation and TVS destruction.

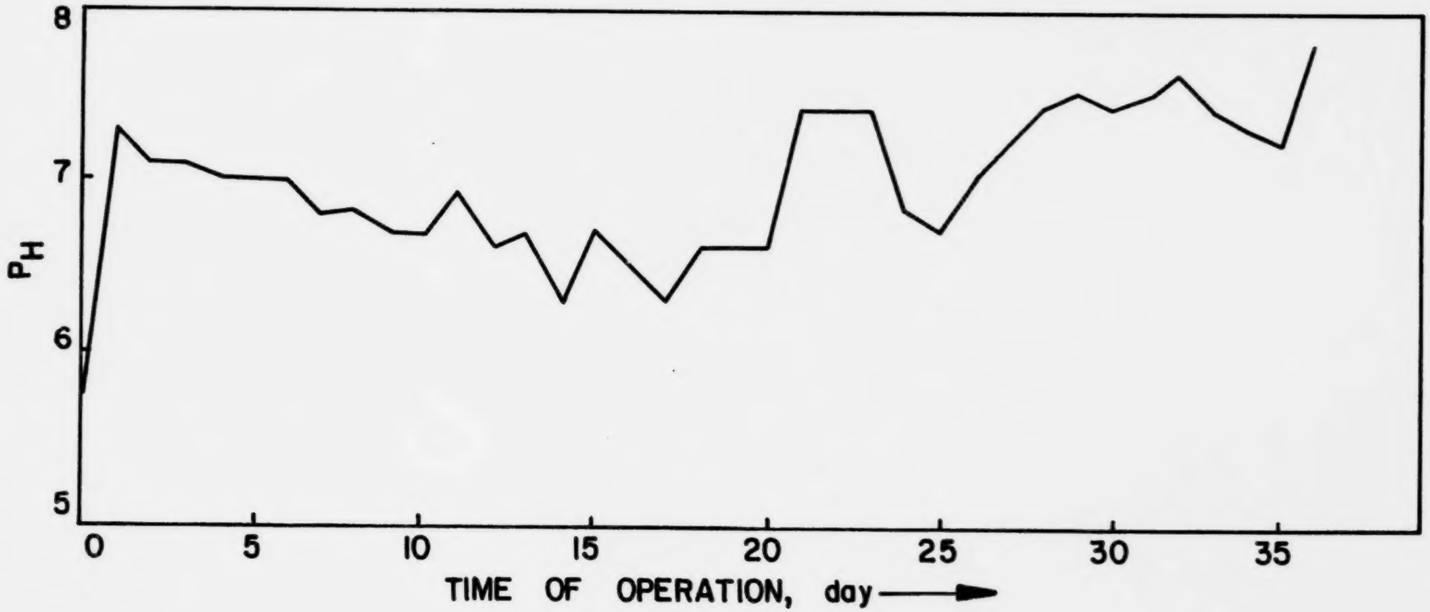

Figure 48. Relationship of pH with time of operation, Biodegradability Study No. 4, Unit No. 1.

to the appropriate steady state conditions with respect to matching dates of operation. However, it should be noted that the batch decay studies were conducted on grab samples, whereas the full scale system received constantly varying sludge feed.

The biodegradability studies were completed at the end of August, 1978; however, steady state (full scale) operation continued into November, 1978. The average results of biodegradability studies 1 through 4 were also used for the full scale performance data collected during this time period in terms of biodegradability because these results were collected over a long time period which included a similar period (September-November) in 1977.

## SYSTEM PERFORMANCE CHARACTERISTICS

The following section presents the results of both the full and bench scale aerobic digesters with respect to system performance as determined by various parameters including temperature, substrate removal efficiencies, reaction rate coefficients, and pathogen destruction. Both the bench and full scale units were operated over a two-year time frame and the feed substrate for both systems was fed as it came from the gravity thickeners. This sludge was variable with respect to concentration, biodegradable fraction, and rate of biodegradability (as exhibited in the biodegradability studies); therefore, it is difficult to make close comparisons between the two reactors in the full scale study, and between the bench and full scale studies.

### Start Up and General Operation

The following information on high temperature sludge treatment was obtained over a period of time including the coldest winter on record in Central New York. Throughout the study the unit was shut down and started for a number of minor modifications, e.g., feed pump overhaul, reactor radius addition, and pipe insulation. It was interesting to note that start up always occurred immediately and rapidly, regardless of other variables such as sludge or air temperature. No special sludge seed was used in the full scale or laboratory reactors. Typical summer and winter start up characteristics are shown in Figure 49. After start up(s) and throughout the study, the reactor temperature was always in the thermophilic zone, i.e., greater than 43°C. It was often above 50°C and reached a maximum of 65°C in early winter of 1978. The pH of the system was fairly constant, varying between 7.0 and 8.0. The system was operated as a continuous feed (hourly, with the exception of the 10-day hydraulic retention time study which was fed every eighth hour) complete-mix, no recycle digester. Therefore, the hydraulic retention time (HRT) was theoretically equal to the solids retention time (SRT) with the loading conditions being altered by adjustment of the sludge (feed) flow rate. Each loading condition was defined in terms of hydraulic loading rate (HRT, days), organic loading rate (LR, kg/$m^3$-day), organic removal rate (RR, kg/$m^3$-day), organic treatment efficiency (TE, % initial removed or oxidized), reaction rate coefficient ($K_T$, $day^{-1}$), autoheated reactor temperature ($T_R$, °C), and aerator performance.

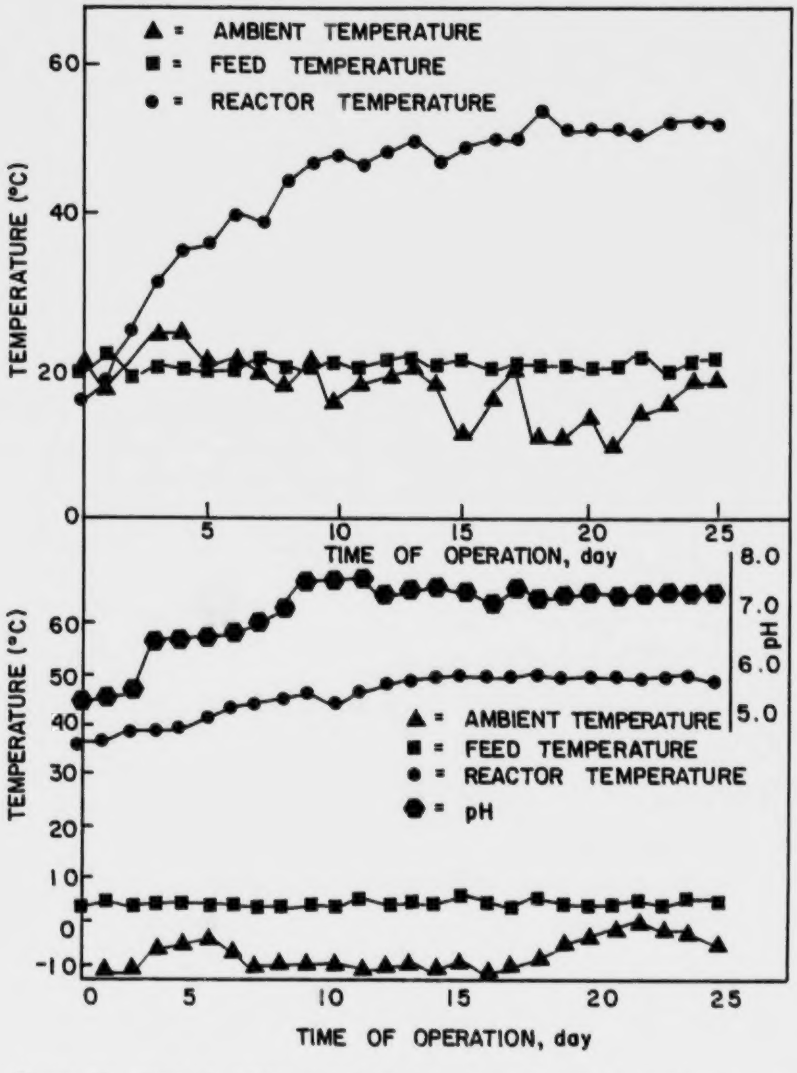

Figure 49. Rate of autoheating during start-up of aerobic sludge digestion during summer and winter conditions.

Each loading condition at the bench scale was controlled and defined as above for the full scale system with the exception of aerator performance. The bench reactors were not autoheated and usually had temperature controlled at about 50°C.

## Steady State Performance

The full scale and bench scale units were operated under controlled conditions for extended periods of time to achieve steady state results as well as to document the practical operational problems (see subsequent Section) that might be expected with this process. Steady state conditions were test conditions and operating results that remained constant over a significant period of time. In most cases this referred to a testing condition at a constant hydraulic retention time where effluent characteristics, reactor temperature, and other major parameters changed less than fifteen percent from one sampling to the next, separated by a period equal to one hydraulic retention time. It is important to note here that for both conditions studied with the Midland-Frings aerator, only one steady state point was selected as representative of typical performance data. That point, in each case, was gathered when the system had experienced continuous operation, as described below, without interruption. The other measurements of performance data, as reported in Appendix D, were gathered when the system was marginally operational due to grit accumulations in the reactor or to sporadic performance of the aerator. Steady state testing for all conditions occurred after operation for a minimum period equal to three hydraulic retention periods after changing major parameters.

Thirty full scale system steady state data points were collected (a total of 38 conditions were tested) in an effort to define five loading conditions with the two aerators. Twenty-one steady state data points with the bench scale system were collected (a total of 58 data conditions were tested) in an effort to define four loading conditions. The average performance data for the nine conditions are presented in Table 16. Appendix D contains all fifty-one data points. A complete record of the full scale reactor, ambient, and feed sludge temperature, as well as the corresponding loading rates studied, is presented in Figure 50.

Figures 51 through 56 represent a summary of the substrate (COD) removal kinetics as observed for the full scale system using both aerators, and for the bench scale. As the organic loading rate was increased (by increasing HRT) the organic removal rate also increased due to the increased availability of biodegradable substrate. The maximum removal rate was 6.53 kg COD/$m^3$-day as observed for the full scale system using the Midland Frings aerator and 4.53 kg COD/$m^3$-day with the centri-rator. The bench scale systems exhibited a maximum removal rate of 5.13 kg COD $m^3$-day.

The relationships between organic loading rates, the hydraulic retention time, and fraction of available organics that were stabilized are shown in Figure 52. Half of the total biodegradable organics was oxidized at a 4-day HRT in most systems, and up to 77 percent was oxidized in the long term tests with the Midland-Frings unit. The data indicated that a hydraulic retention time of little more than 2.7 days resulted in BCOD destruction of more than

TABLE 16. PERFORMANCE DATA AS AVERAGED FOR STEADY STATE CONDITIONS FOR BOTH FULL AND BENCH SCALE AEROBIC DIGESTERS

| | Bench | Full (C)* | Full (M)* | Full (C)* | Bench | Bench | Bench | Full (C)* | Full (M)* |
|---|---|---|---|---|---|---|---|---|---|
| $\Theta_H$(day) | 2.86 | 3.20 | 3.49 | 4.56 | 5.00 | 5.00 | 7.00 | 7.74 | 11.0 |
| TCOD Loading Rate† | 13.2 | 17.7 | 14.6 | 12.1 | 12.1 | 11.1 | 5.63 | 5.18 | 4.36 |
| TS Loading Rate† | 12.6 | 17.9 | 15.6 | 12.0 | 13.5 | 10.2 | 5.60 | 4.40 | 3.49 |
| TVS Loading Rate† | 7.86 | 11.0 | 9.75 | 7.48 | 7.86 | 6.26 | 3.71 | 2.78 | 2.53 |
| TCOD Removal Rate† | 5.13 | 4.53 | 6.53 | 3.73 | 4.05 | 4.14 | 2.70 | 2.35 | 2.51 |
| TS Removal Rate† | 1.54 | 2.74 | 8.30 | 2.58 | 3.71 | 2.33 | 2.10 | 1.07 | 1.43 |
| TVS Removal Rate† | 2.59 | 2.80 | 4.89 | 2.35 | 2.65 | 1.67 | 1.79 | 1.00 | 1.25 |
| TCOD Treat. Eff. (%) | 38.8 | 25.5 | 44.9 | 30.8 | 33.6 | 37.3 | 48.0 | 45.5 | 57.5 |
| TS Treat. Eff. (%) | 29.4 | 14.9 | 50.2 | 21.3 | 27.2 | 22.9 | 37.5 | 24.0 | 41.1 |
| TVS Treat. Eff (%) | 37.7 | 25.2 | 53.3 | 31.3 | 33.4 | 39.1 | 48.5 | 34.2 | 49.4 |
| TKN Feed (mg/ℓ) | - - | 2268 | - - | - - | 3010 | - - | - - | 1279 | 1435 |
| TKN Effluent (mg/ℓ) | - - | 2023 | - - | - - | 2156 | - - | - - | 968 | 1323 |
| $NH_3$-N Feed (mg/ℓ) | - - | 188 | - - | - - | 42 | - - | - - | 103 | 105 |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | 62 | - - | - - | 235 | - - |

(Continued)

TABLE 16 (continued)

| | Bench | Full (C)* | Full (M)* | Full (C)* | Bench | Bench | Bench | Full (C)* | Full (M)* |
|---|---|---|---|---|---|---|---|---|---|
| pH Feed | 6.5 | 5.8 | 5.4 | 5.8 | 6.0 | 5.8 | 5.7 | 5.5 | 5.7 |
| pH Effluent | 7.0 | 6.8 | 6.8 | 6.9 | 7.6 | 6.7 | 7.1 | 7.7 | 7.1 |
| Temp. Reactor‡ | 55.0 | 50.0 | 61.8 | 48.0 | 51.5 | 47.7 | 54.5 | 46.7 | 62.3 |
| Temp. Ambient‡ | 15.0 | 14.5 | 17.7 | 1.0 | 15.0 | 15.0 | 15.0 | 19.0 | 6.8 |
| Temp. Feed‡ | - - | 11.8 | 20.0 | 6.3 | - - | - - | - - | 21.6 | 11.9 |
| Capillary Suction Time Feed (secs) | - - | 500 | 75 | 408 | 69 | - - | - - | - - | 44 |
| Capillary Suction Time Effluent (secs) | - - | 1095 | 553 | 874 | 964 | - - | - - | - - | 145 |
| Airflow (ℓ/sec) | 2.0 | 26.0 | 39.6 | 25.5 | 2.8 | 2.0 | 1.9 | 25.2 | 40.0 |
| Oxygen Transfer Eff. (%) | - - | 22.9 | - - | - - | - - | - - | - - | 8.67 | - - |
| Oxygen Transfer Rate (kg/kW-hr) | - - | 1.94 | - - | - - | - - | - - | - - | 0.59 | - - |
| Oxygen Transfer Eff. (COD Basis) (%) | - - | 18.9 | 22.1 | 18.3 | - - | - - | - - | 13.8 | 7.9 |

(continued)

TABLE 16 (continued)

| | Bench | Full (C)* | Full (M)* | Full (C)* | Bench | Bench | Bench | Full (C)* | Full (M)* |
|---|---|---|---|---|---|---|---|---|---|
| Oxygen Transfer Rate (COD Basis) (kg/kW-hr) | - - | 1.62 | 1.80 | 1.22 | - - | - - | - - | 0.87 | 0.66 |
| No. of points used in Average (N) | 4 | 4 | 1 | 4 | 3 | 3 | 2 | 3 | 1 |

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[*] Identification: (C) = centri-rator; (M) = Midland-Frings.

[†] $kg/m^3$-day.

[‡] °C.

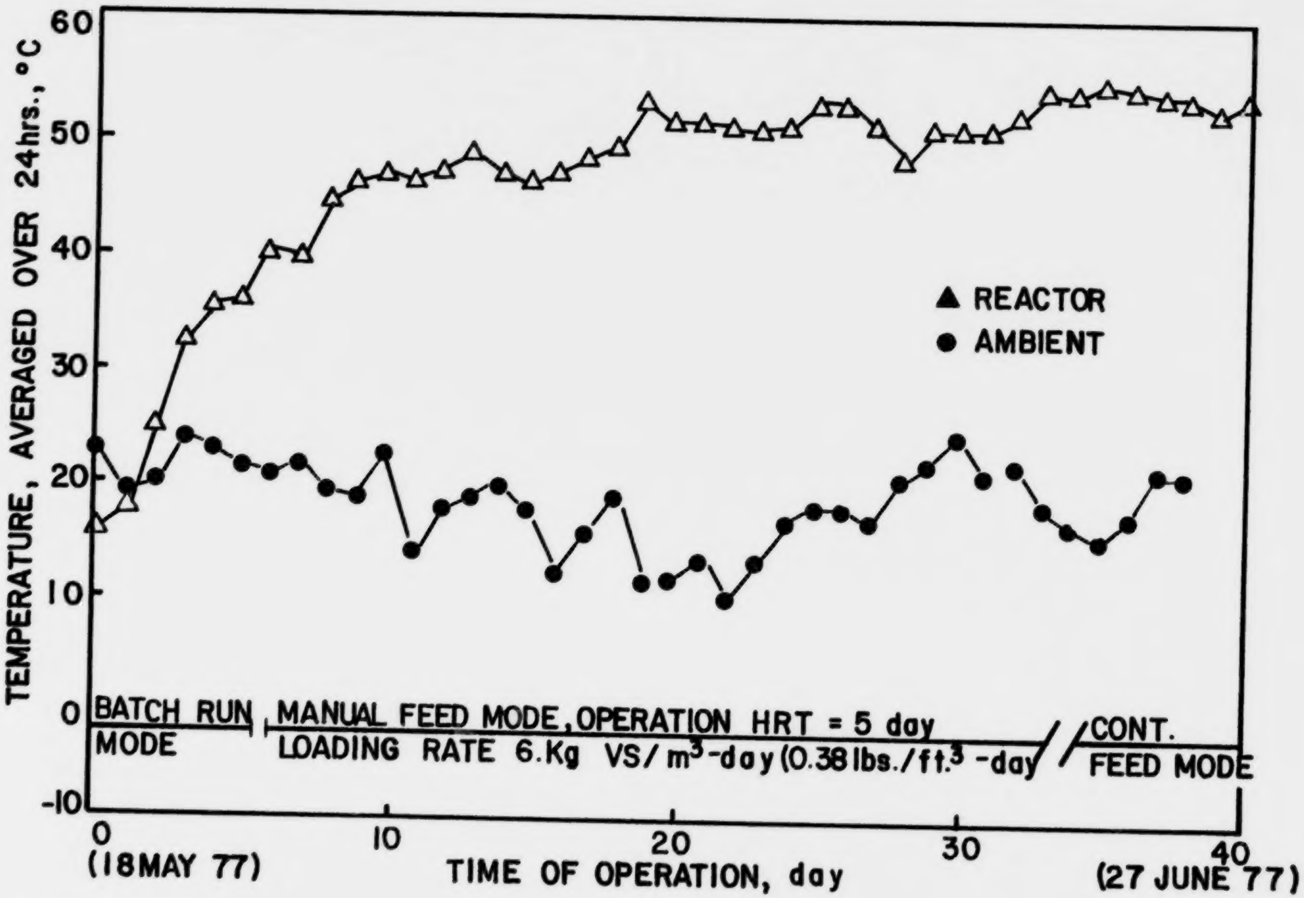


Figure 50. Reactor (autoheated), ambient, and feed temperatures with corresponding loading rates and operational modes for the duration of the full scale study (May 19, 1977 - November 24, 1978).

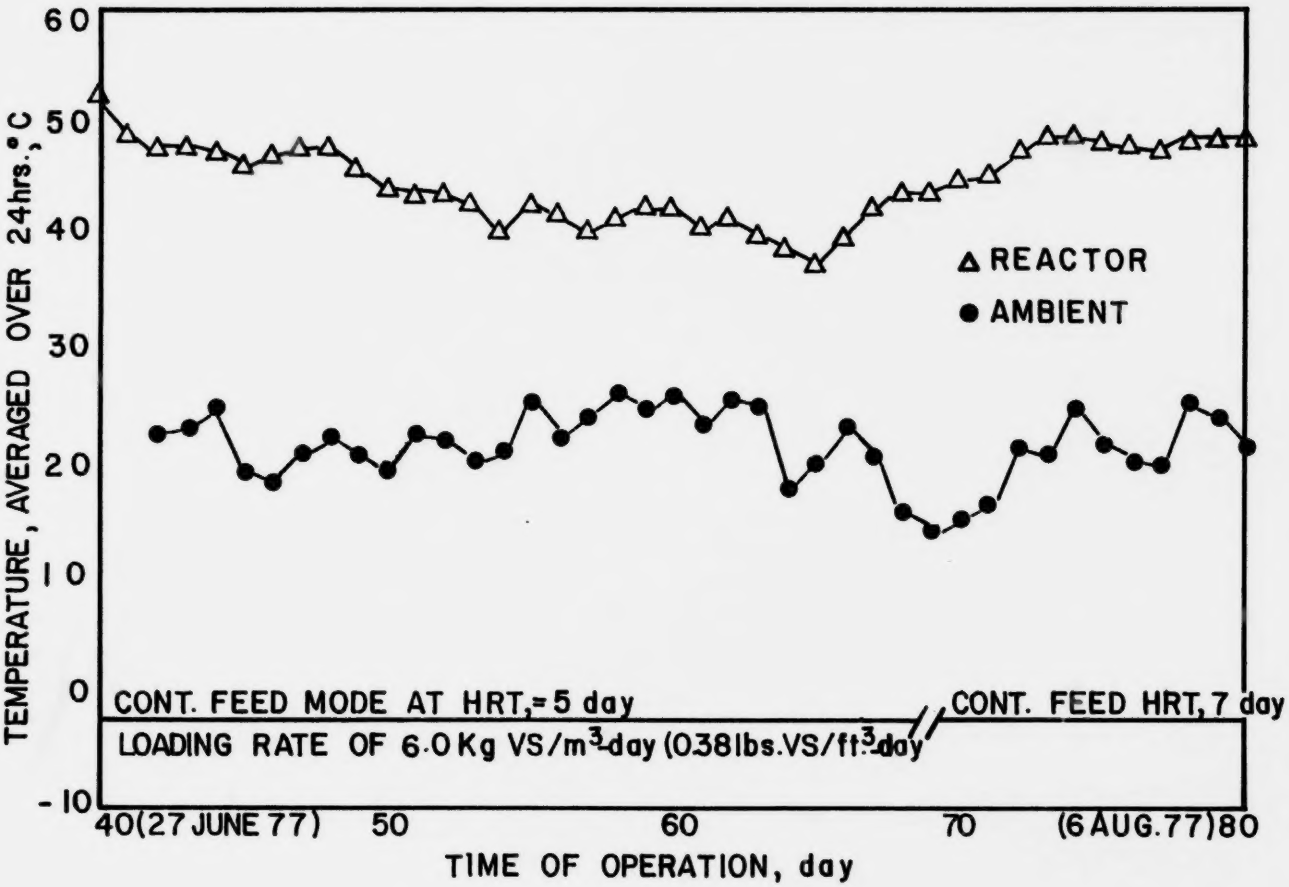
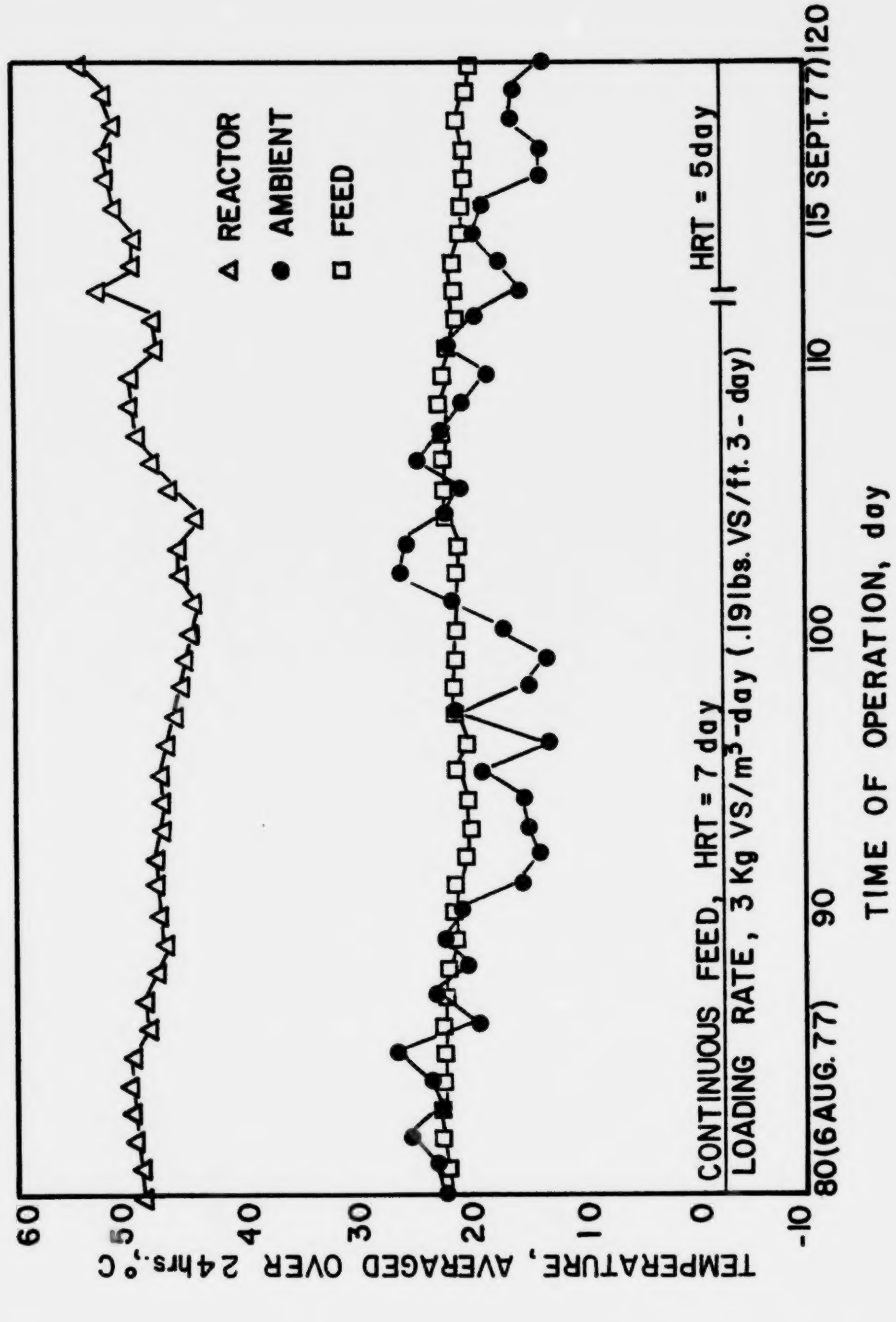


Figure 50. (continued)



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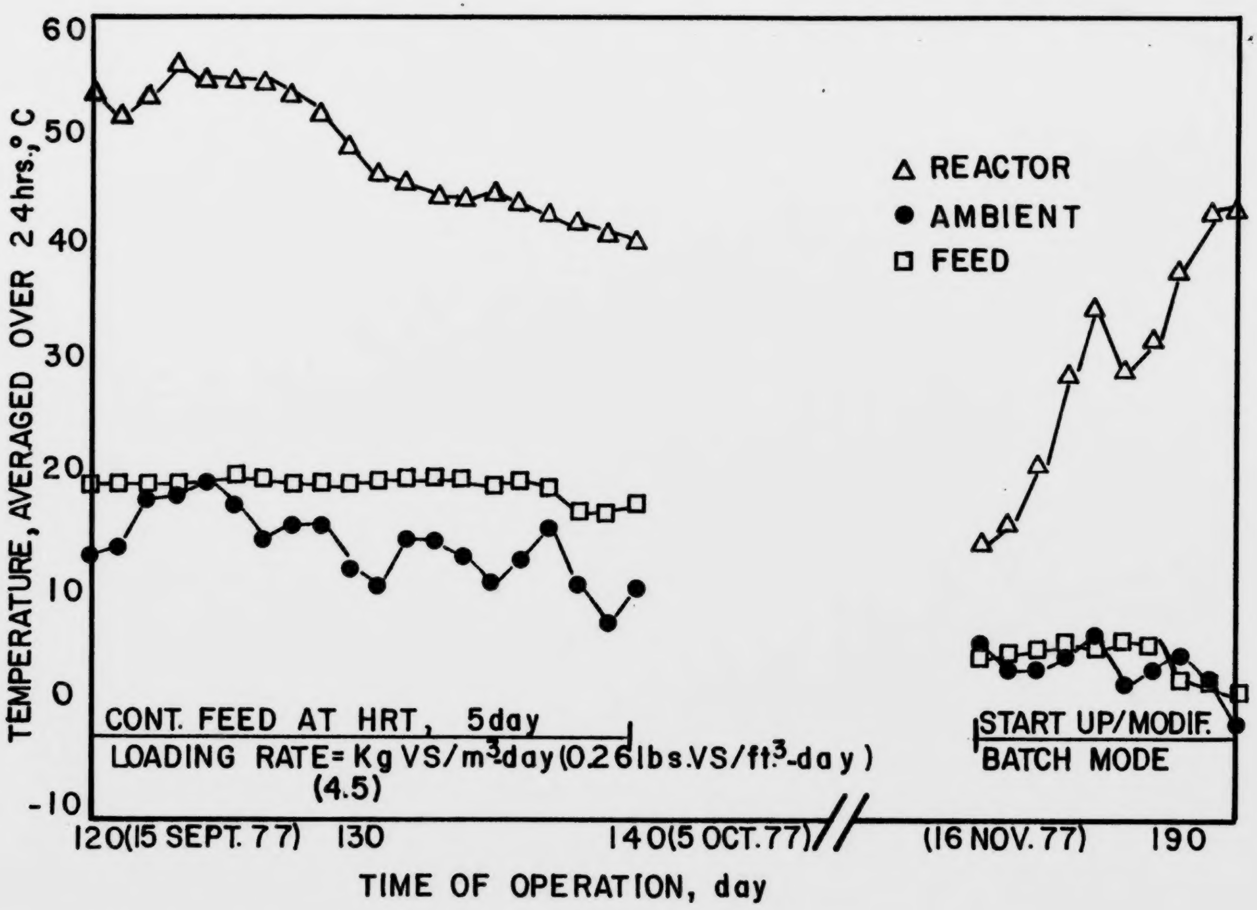


Figure 50. (continued)

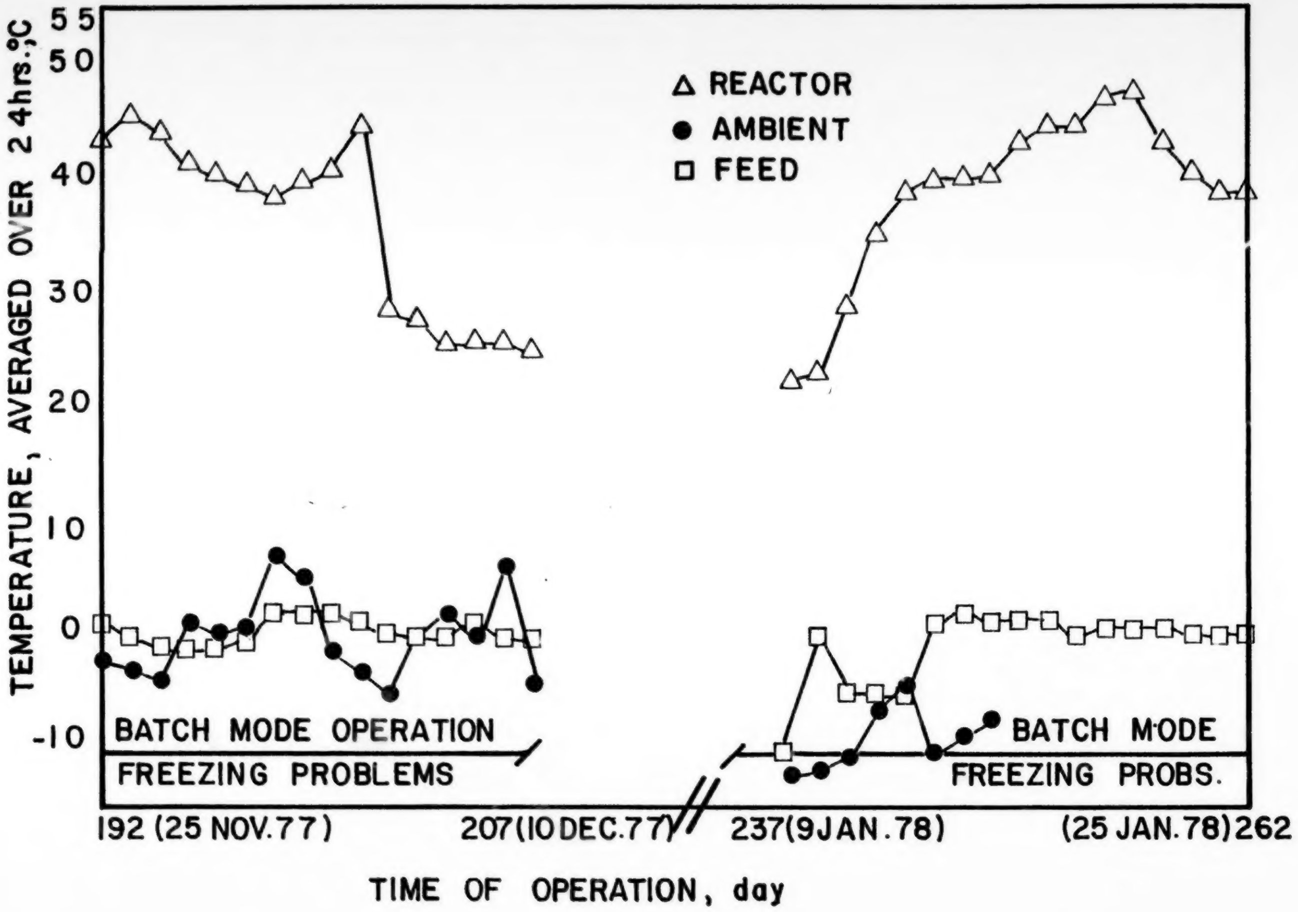


Figure 50. (continued)

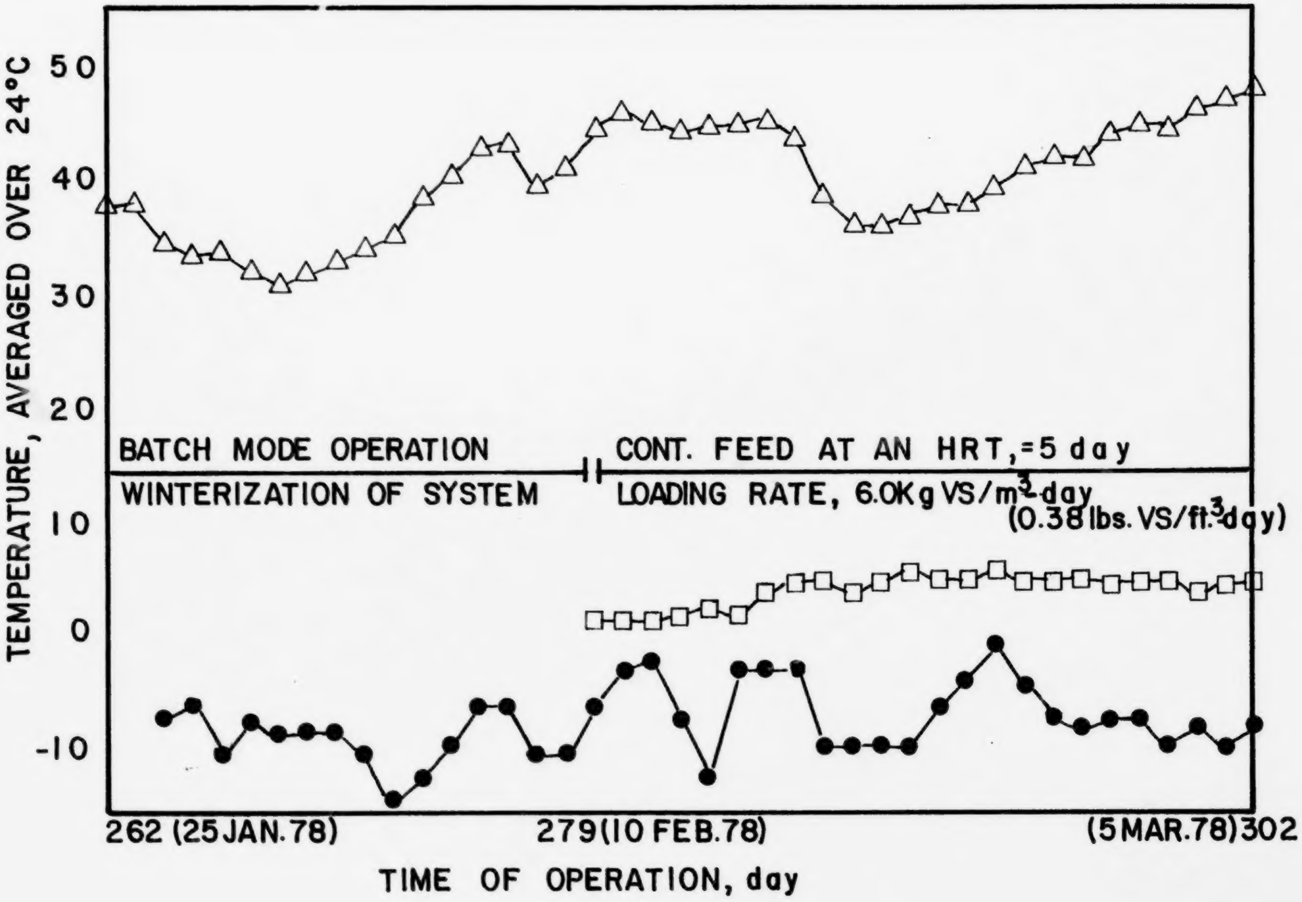


Figure 50. (continued)

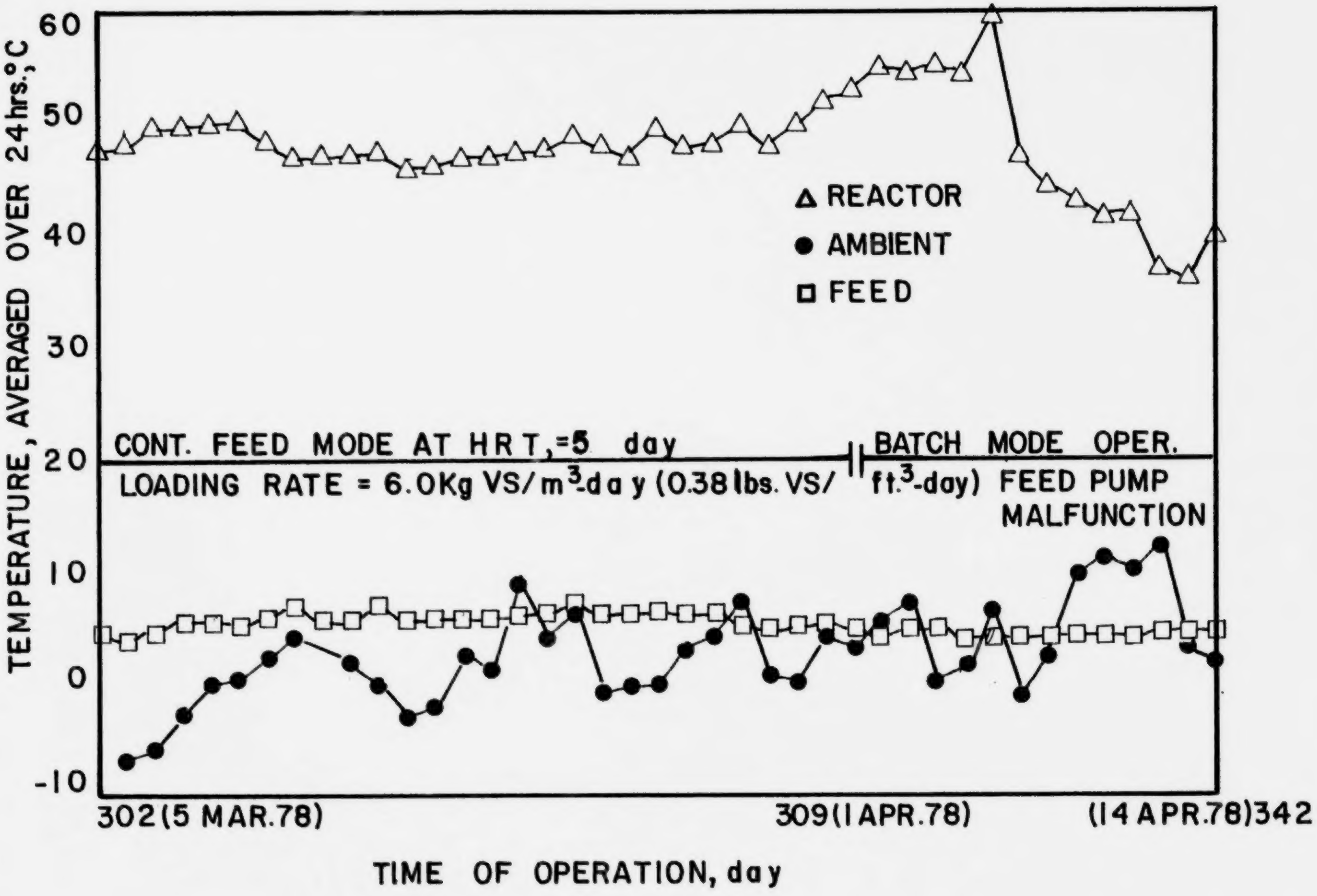


Figure 50. (continued)

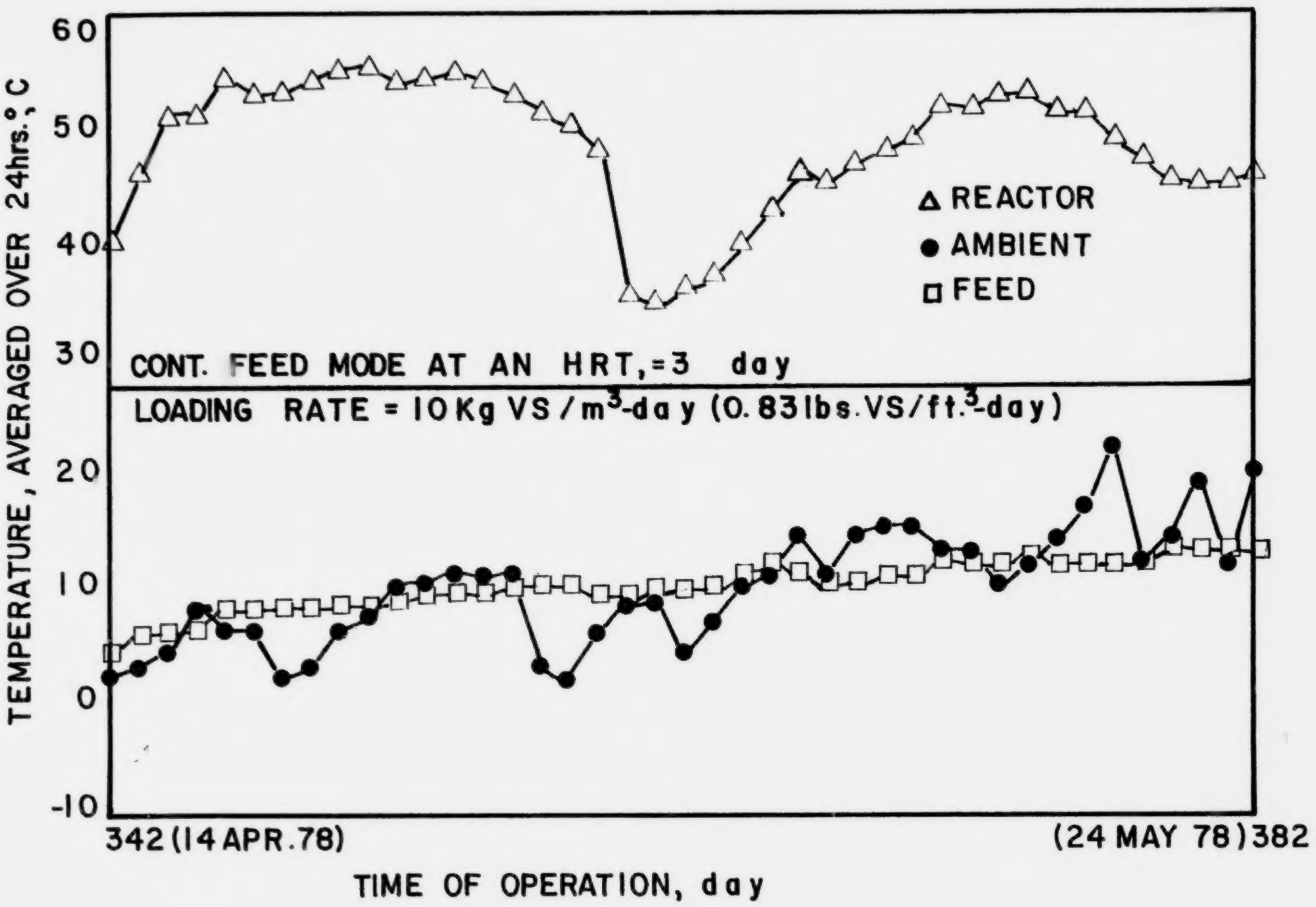


Figure 50. (continued)

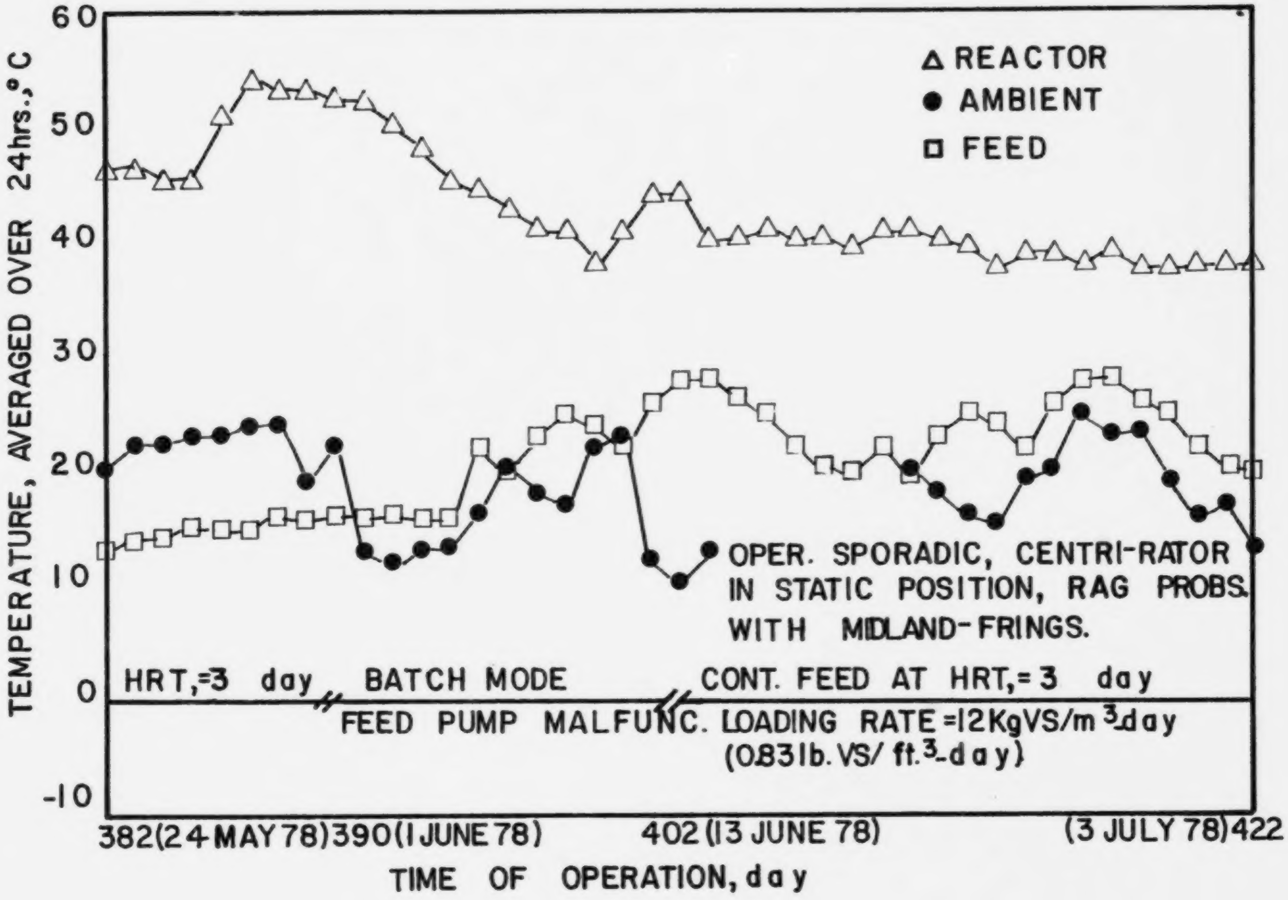


Figure 50. (continued)

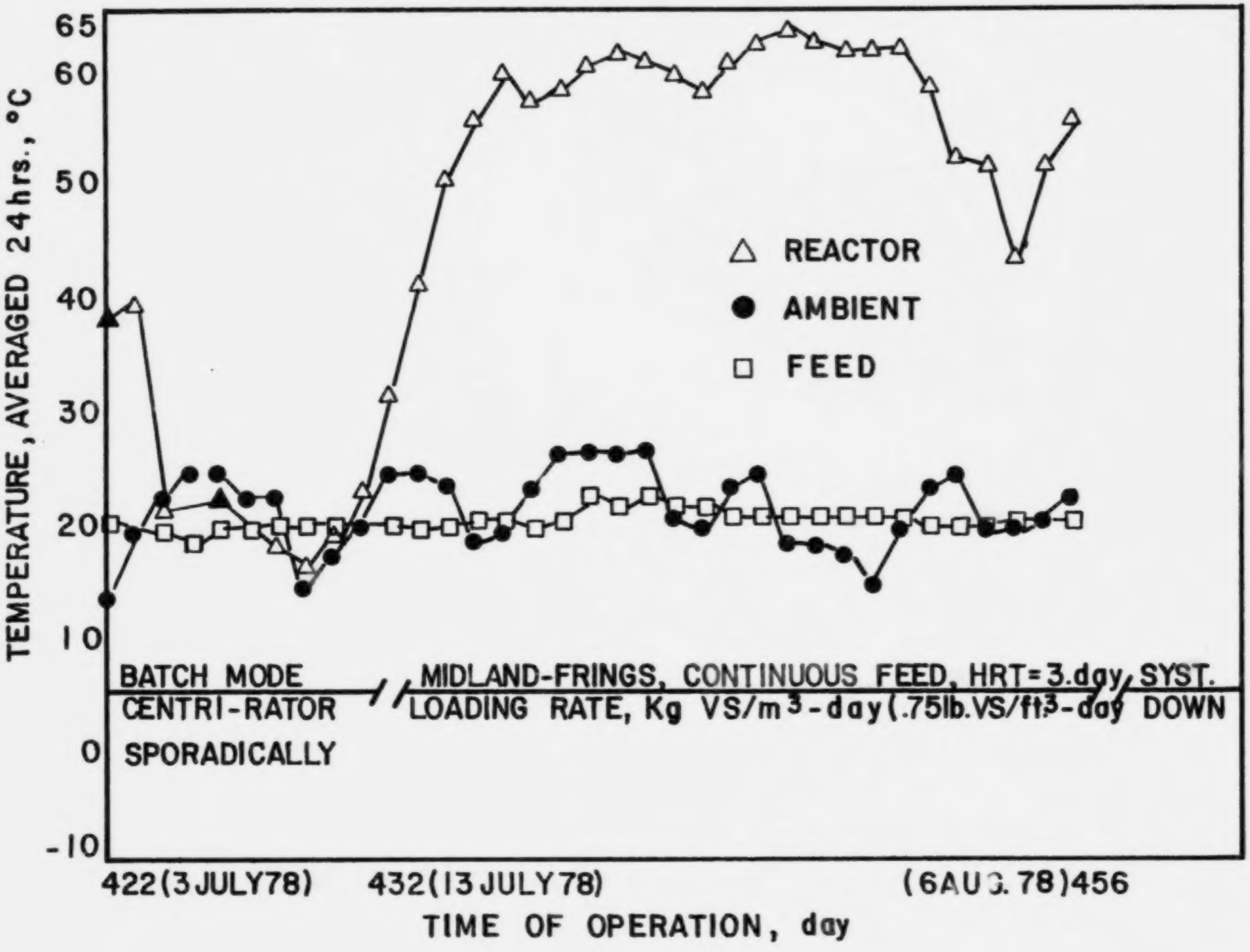


Figure 50. (continued)

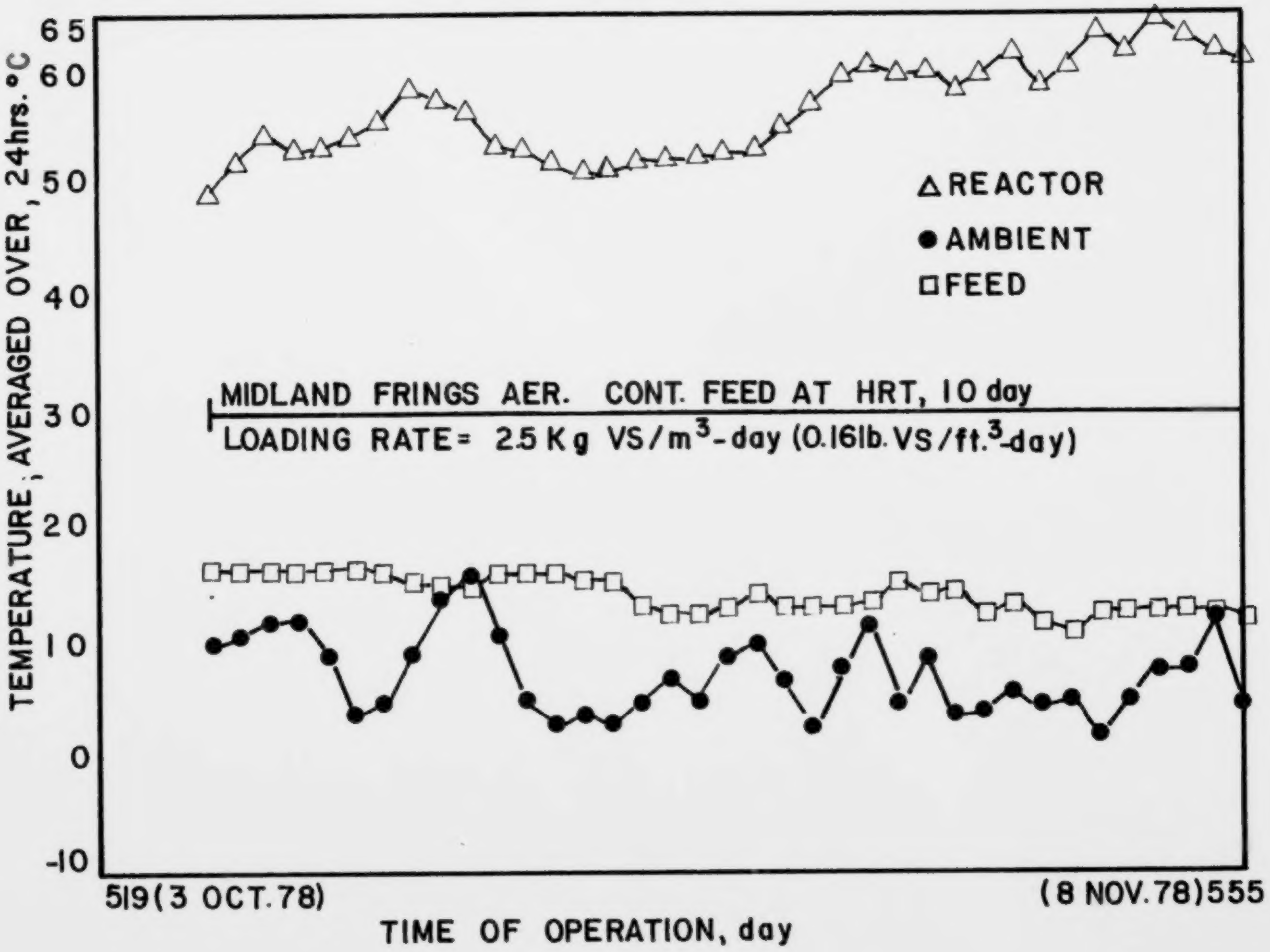


Figure 50. (continued)

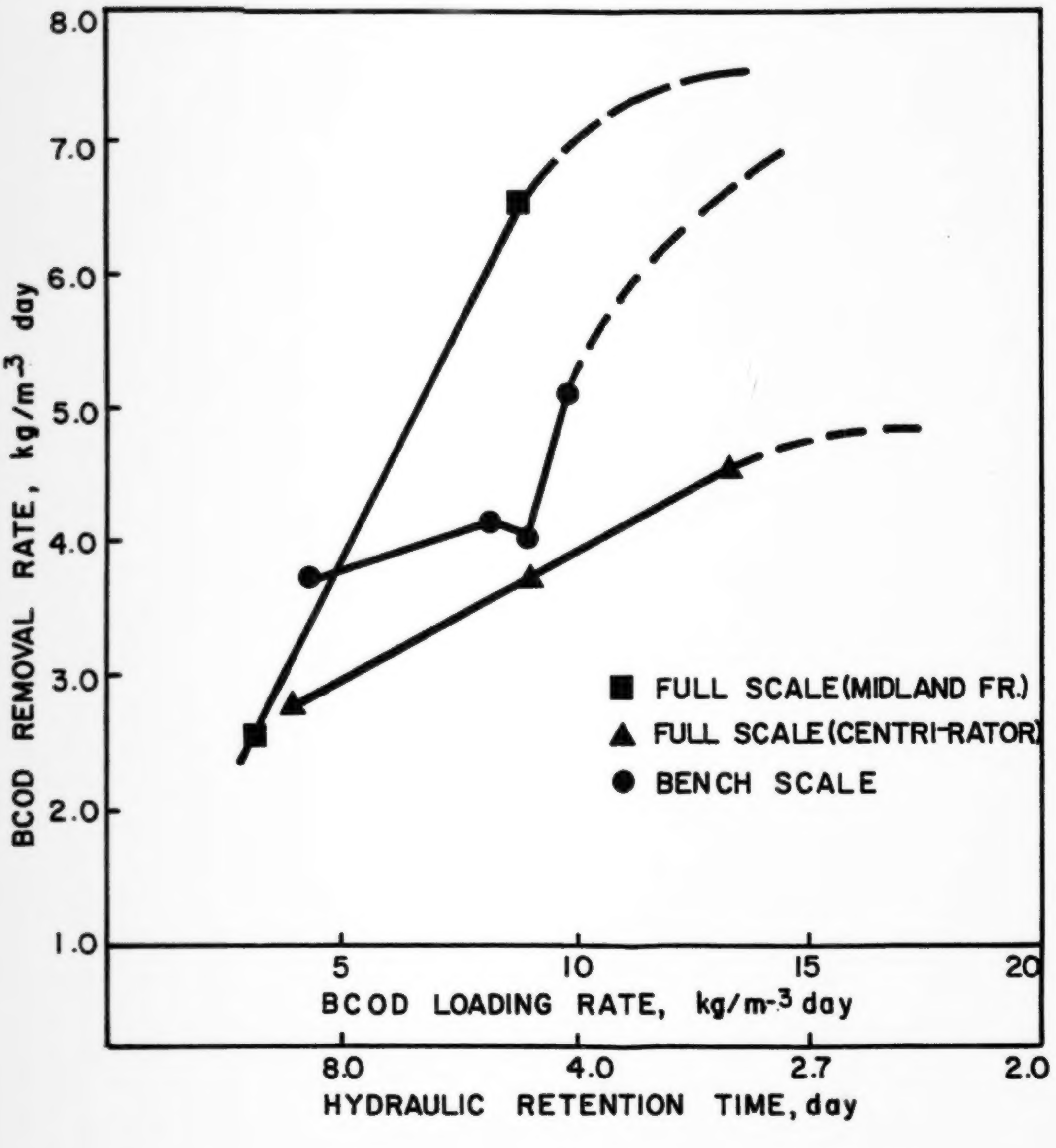


Figure 51. Relationship observed for removal rate (BCOD) versus loading rate (BCOD) and hydraulic retention time (HRT)$^+$ for bench and full scale systems at steady state conditions. ($^+$HRT based on constant feed concentration of 40 gm BCOD/ℓ.)

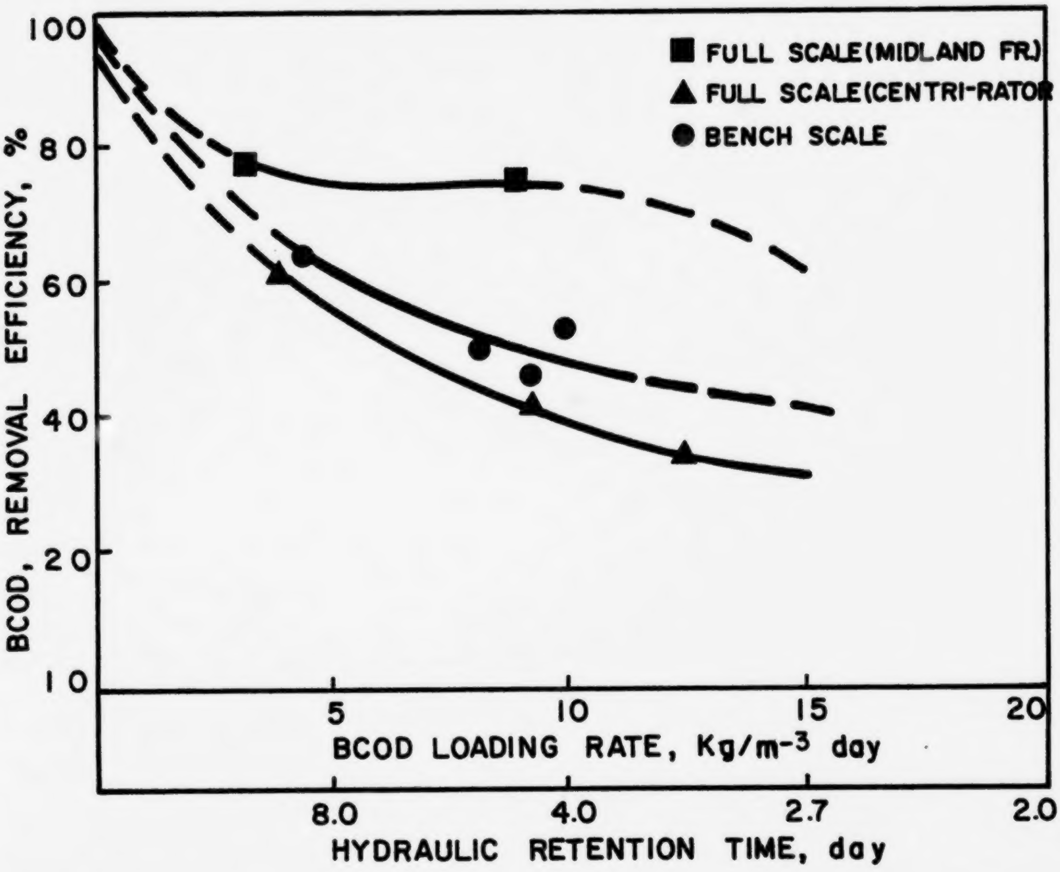


Figure 52. Relationship observed for removal efficiency (BCOD) versus loading rate (BCOD) and hydraulic retention time (HRT)$^+$ for bench and full scale systems at steady state conditions. ($^+$HRT based on constant feed concentration of 40 gm BCOD/ℓ.)

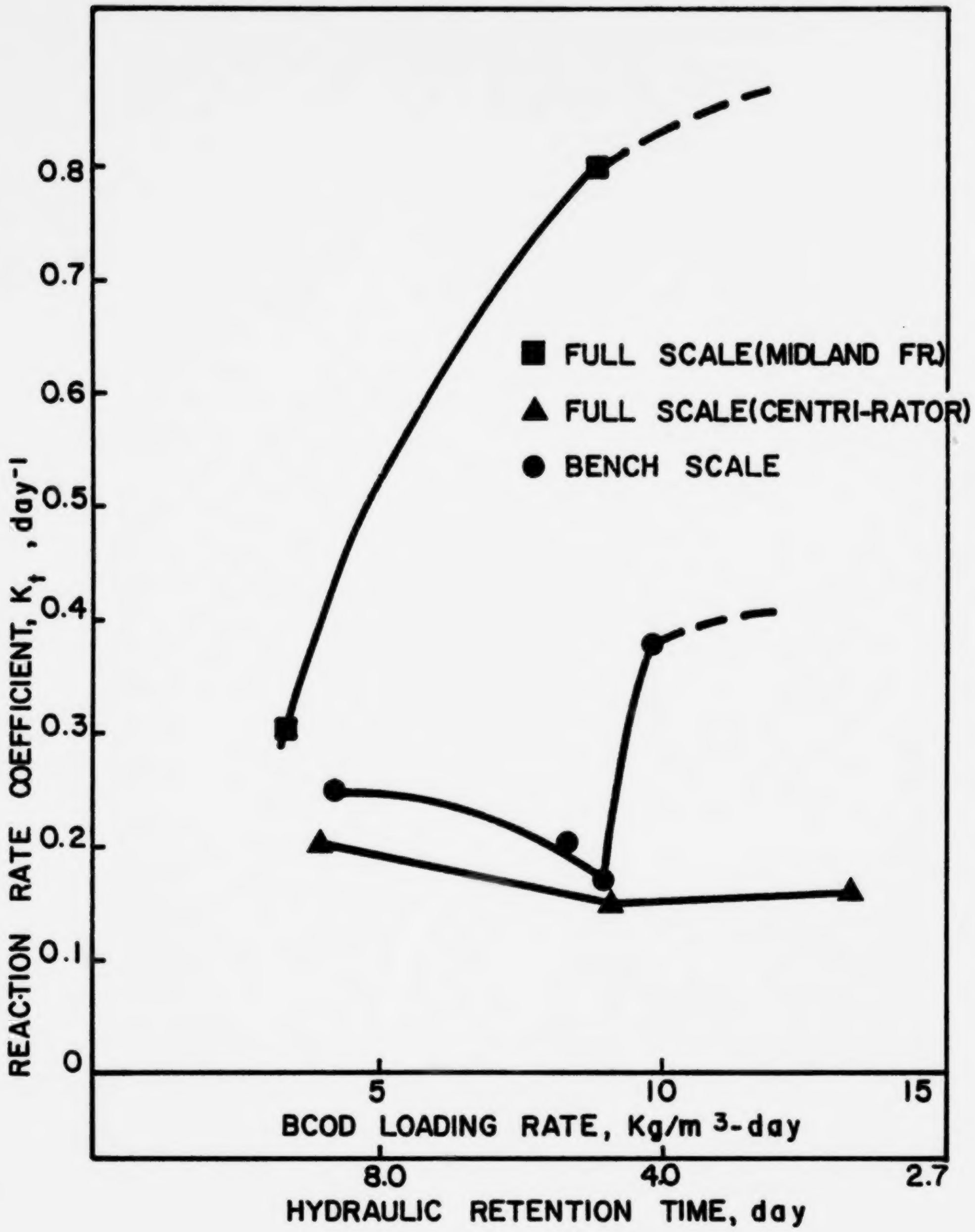


Figure 53. Reaction rate coefficient as related to organic loading rate (BCOD) for both bench scale and full scale studies (HRT based on constant feed concentration of 40 gm BCOD/ℓ.)

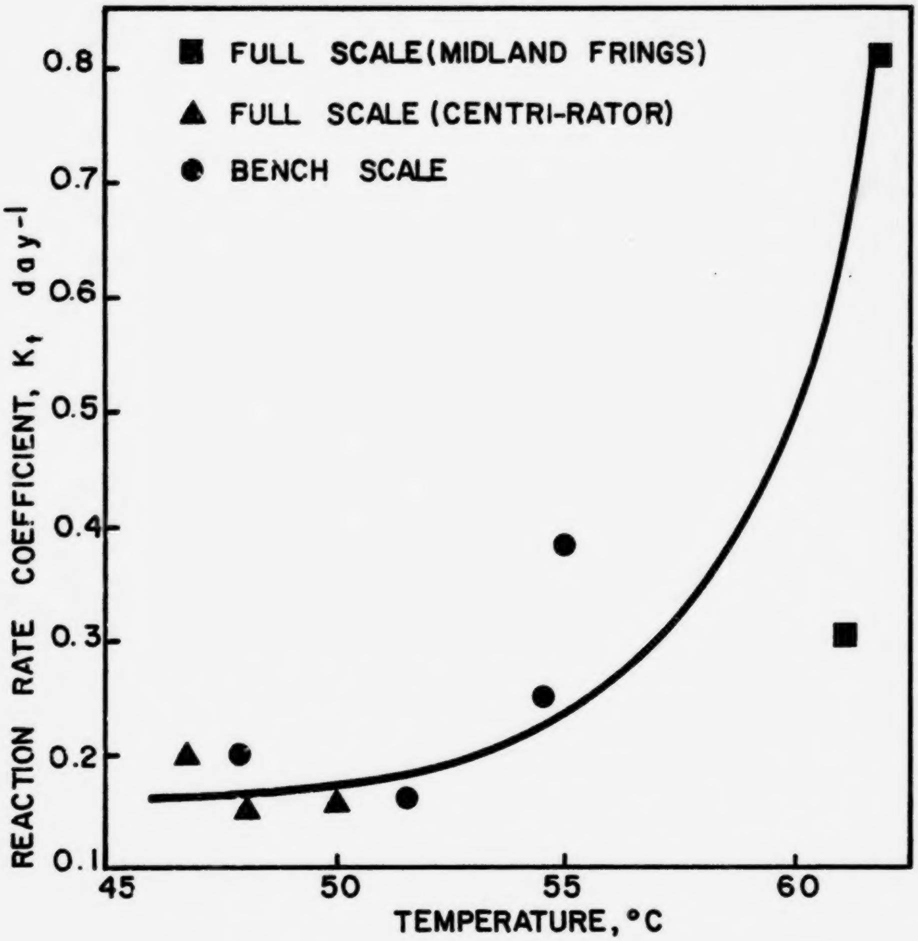


Figure 54. Reactor rate coefficient ($K_T$) as a function of temperature (T) for full scale and bench scale reactors.

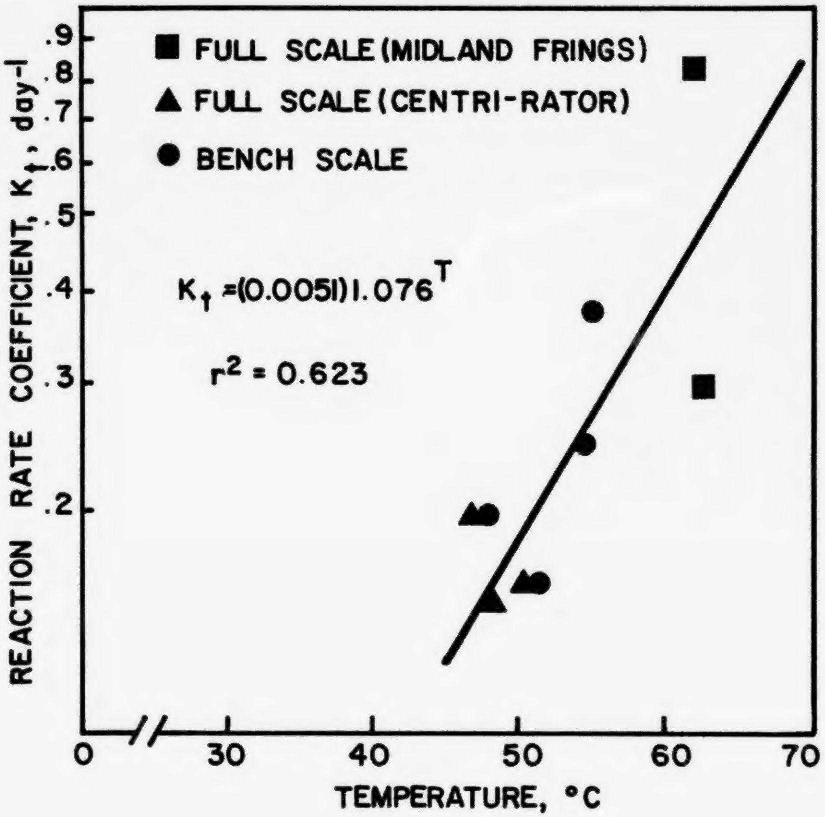


Figure 55. Exponential relationship observed for the reaction rate coefficient ($K_T$) as a function of temperature.

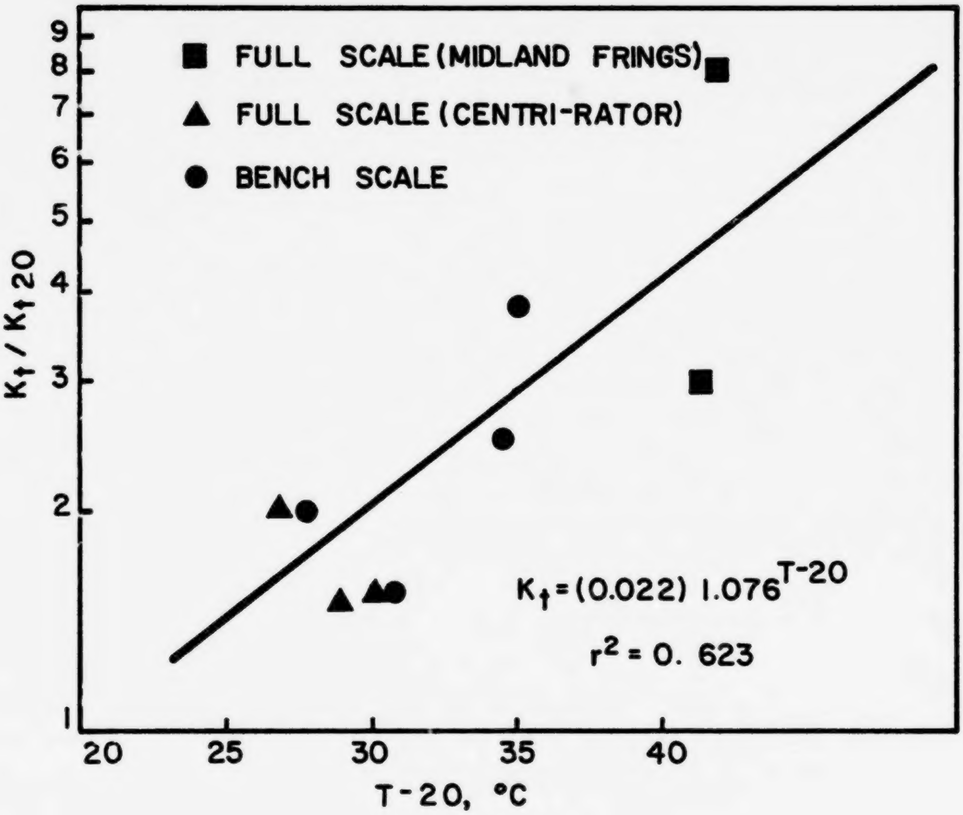


Figure 56. Effect of temperature on the reaction rate coefficient ($K_T$) (compared to $K_T$ at 20°C) when $K_{T\ 20} = 0.10\ \text{day}^{-1}$ and biodegradable substrate ≙ 75% of total.

35 percent. Organic loading rates higher than that caused by the hydraulic retention time of about three days exceeded the oxygen transfer capability of the aerators.

The removal rate coefficient, $K_T$, increased with loading and was highest with the Midland-Frings aerator at 0.8 day$^{-1}$, followed by the bench scale and then the full scale using the centri-rator (see Figure 53). The increase of $K_T$ with loading was probably a function of higher dissolved oxygen levels and increased temperature achieved with the Midland Frings unit. The relationship between $K_T$ and reactor temperature is shown in Figures 54 and 55. These data support earlier studies which showed that $K_T$ was a function of temperature (2,21). Figure 56 presents the temperature function of $K_T$ in a more conventional manner by relating to $K_T$ at 20°C. Both curves illustrate the exponential relationship of $K_T$ and temperature, for those temperatures between 45°C and 65°C. The temperature coefficient, $\theta$, developed from a linear regression analysis of these data, was found to be about 1.08 in both curves. This value of $\theta$ is higher than values previously reported of 1.023 which reflects a doubling in biological activity rates. The depression of RR and $K_T$ in the bench scale system and full scale system with the centri-rator at intermediate loadings was most likely due to low dissolved oxygen levels, i.e., oxygen limited reactions. A complete discussion on the performance variables and their interplay will be presented in a subsequent section.

Autoheating as a Function of Reactor Loading--
The total quantity of heat generated throughout the study as reflected by reactor temperatures is shown in Figures 57 and 58. The Midland-Frings aerator enabled a maximum temperature development of greater than 55°C over the temperature of the feed sludge. Even at the maximum organic loading rates of 13.3 kg/m$^3$-day, the autoheated temperature increase approached 40°C.

A summary of the heat losses for operation with the Midland-Frings and the centri-rator is given in Figures 59 and 60. Table 17 shows all the heat balance and kinetic parameters for steady state operation. Heat losses to the effluent were generally the greatest individual heat loss from the system (the exception being the Midland-Frings unit at low loadings). The importance of heat conservation through efficient insulation and through aeration with low airflow (and high oxygen transfer efficiency) is shown by the relatively low values of $H_A$ and $H_{LA}$.

As was mentioned earlier, the maximum temperature development occurred with the Midland-Frings at about 77 percent removal of biodegradable organics. This removal efficiency corresponded to an organic loading rate of 3.27 kg BCOD/m$^3$-day and a removal rate of 2.51 kg BCOD/m$^3$-day. The maximum temperature development by the centri-rator occurred at a loading rate of 13.3 kg BCOD/m$^3$-day with a corresponding treatment efficiency of 34 percent (biodegradable basis) and a removal rate of 4.53 kg BCOD/m$^3$-day. The relationship between removal rates and biological heat generation is shown in Figure 61.

The quantity of heat released during microbial respiration, R, is directly related to substrate removal rate. That is to say, R is directly proportional to the amount of TCOD oxidized. The quantity of heat released as related to organic loading rate for each aerator studied is presented in

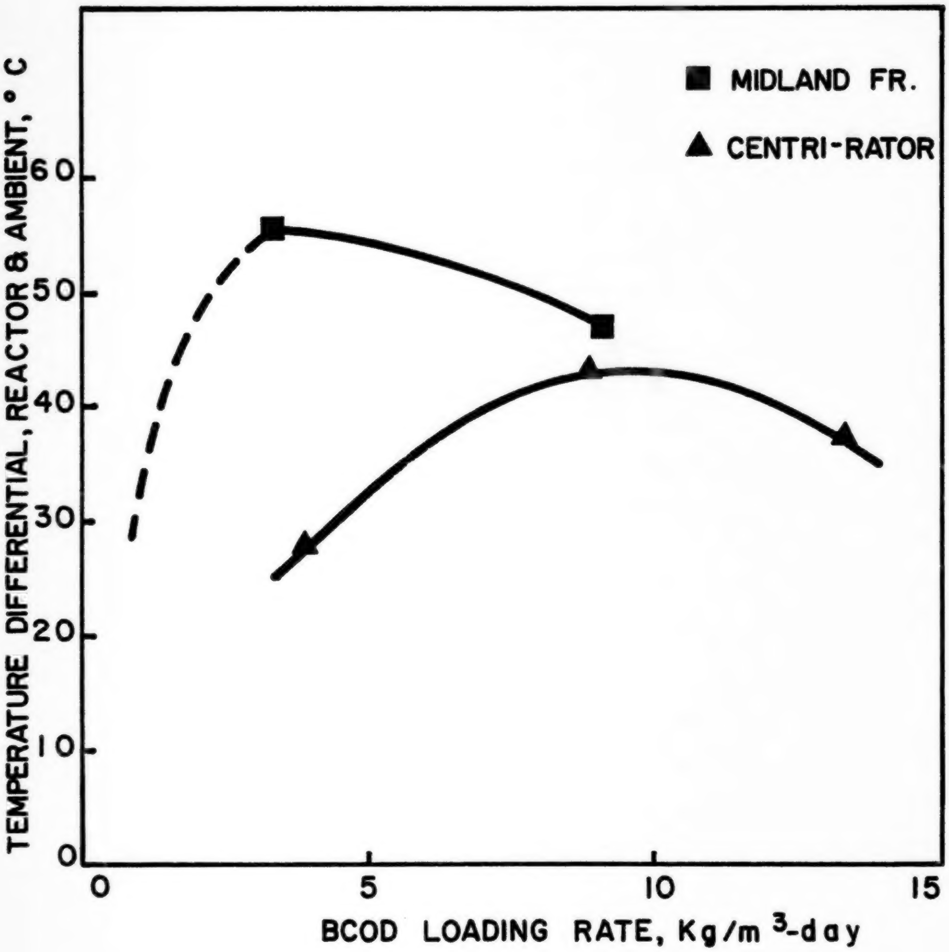


Figure 57. Temperature differential, between the autoheated reactor and ambient temperature observed at steady state conditions, of various biodegradable organic loading rates.

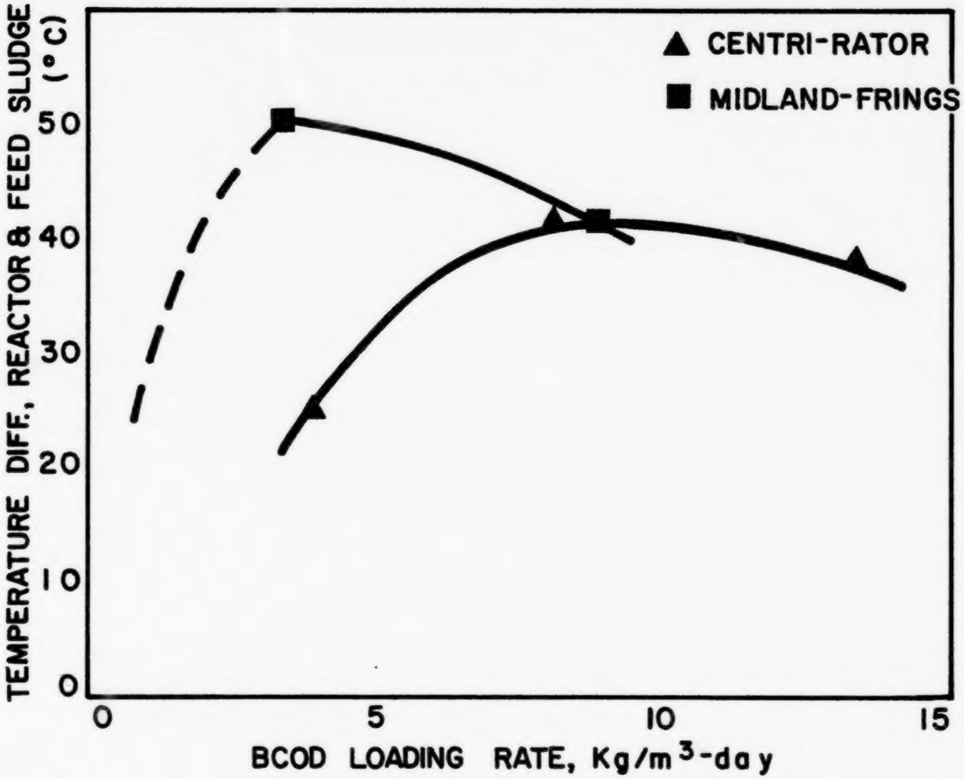


Figure 58. Temperature differential, between the autoheated reactor and feed sludge temperature observed at steady state conditions, at various biodegradable organic loading rates.

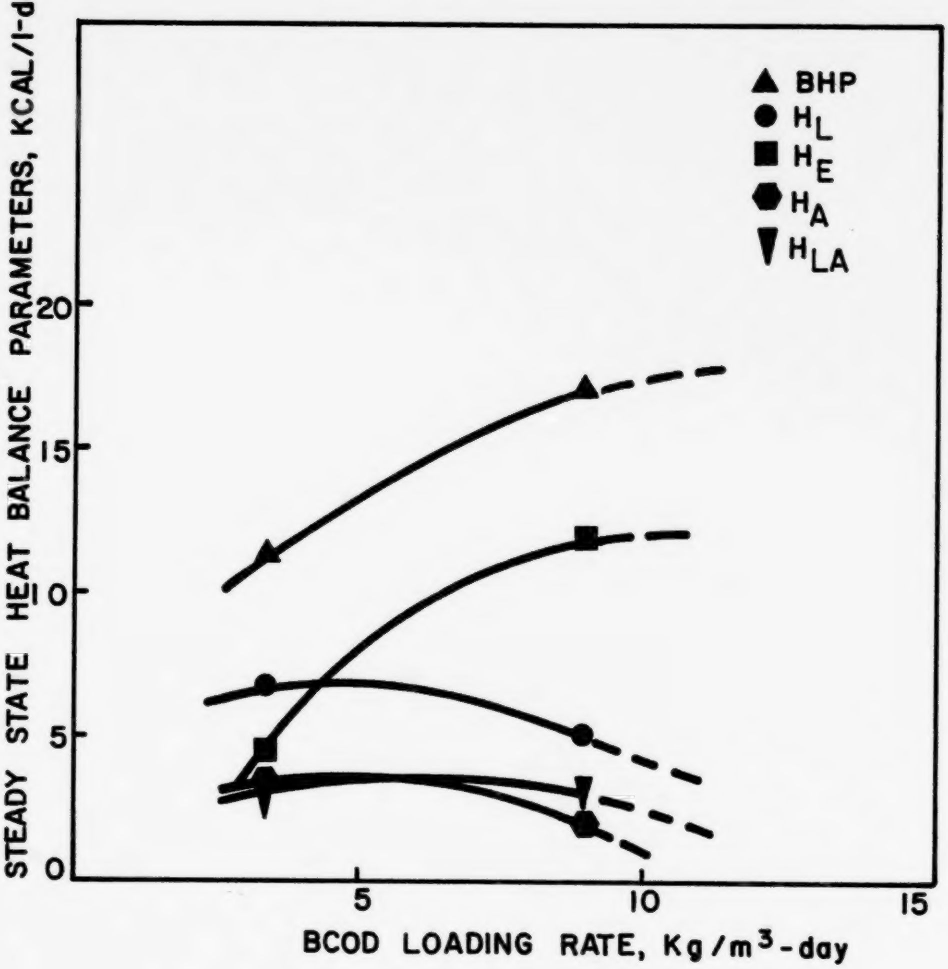


Figure 59. Steady state heat balance parameters as a function of BCOD loading for the Midland-Frings aerator.

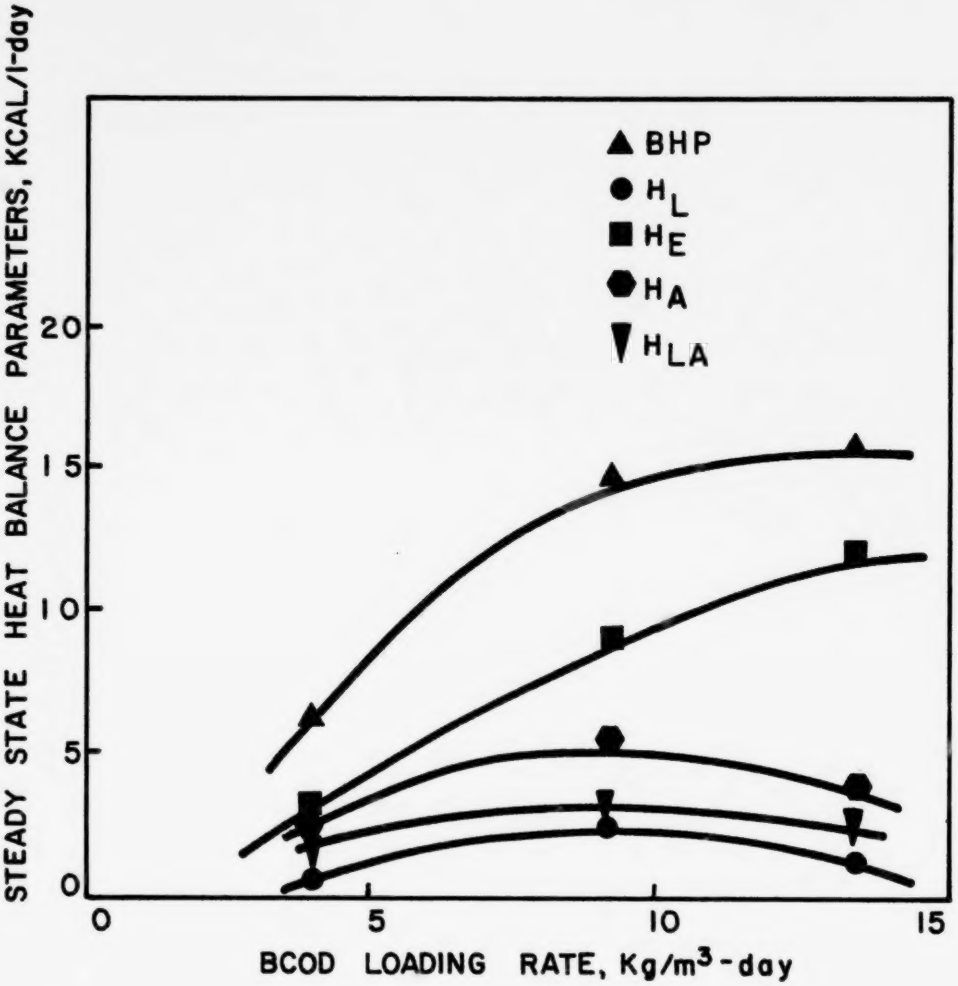


Figure 60. Steady state heat balance parameters as a function of BCOD loading for the centri-rator aerator.

TABLE 17. SUMMARY OF STEADY STATE HEAT BALANCE AND KINETIC PARAMETERS FOR BENCH AND FULL SCALE SYSTFMS.

| ID | Bench | Full (C) | Full (M) | Full (C) | Bench | Bench | Bench | Full (C) | Full (M) |
|---|---|---|---|---|---|---|---|---|---|
| HRT, days | 2.86 | 3.20 | 3.49 | 4.56 | 5.00 | 5.00 | 7.00 | 7.74 | 11.00 |
| BCOD Loading Rate Kg/m$^3$-day | 9.90 | 13.30 | 8.83 | 9.08 | 9.08 | 8.33 | 4.22 | 3.89 | 3.27 |
| BCOD Removal Rate Kg/m$^3$-day | 5.13 | 4.53 | 6.53 | 3.73 | 4.05 | 4.14 | 2.70 | 2.35 | 2.51 |
| BCOD Treatment Efficiency, % | 51.7 | 34.1 | 74.0 | 41.1 | 44.8 | 49.7 | 64.0 | 60.7 | 76.8 |
| Reaction Rate Coefficient $K_T$ day$^{-1}$ | 0.38 | 0.16 | 0.81 | 0.15 | 0.16 | 0.20 | 0.25 | 0.20 | 0.30 |
| $H_L$, Kcal/ℓ-day | -- | 3.88 | 5.11 | 5.55 | -- | -- | -- | 2.82 | 6.86 |
| $H_A$, Kcal/ℓ-day | -- | 1.28 | 2.10 | 2.43 | -- | -- | -- | 0.73 | 3.55 |
| $H_{LA}$, Kcal/ℓ-day | -- | 2.60 | 3.01 | 3.12 | -- | -- | -- | 2.09 | 3.31 |
| $H_E$, Kcal/ℓ-day | -- | 11.9 | 12.0 | 9.14 | -- | -- | -- | 3.24 | 4.58 |
| BHP, Kcal/ℓ-day | -- | 15.8 | 17.1 | 14.7 | -- | -- | -- | 6.06 | 11.4 |
| R, Kcal/gm | -- | 3.49 | 2.62 | 3.94 | -- | -- | -- | 2.58 | 4.54 |

C = Centrirator

M = Midland-Frings

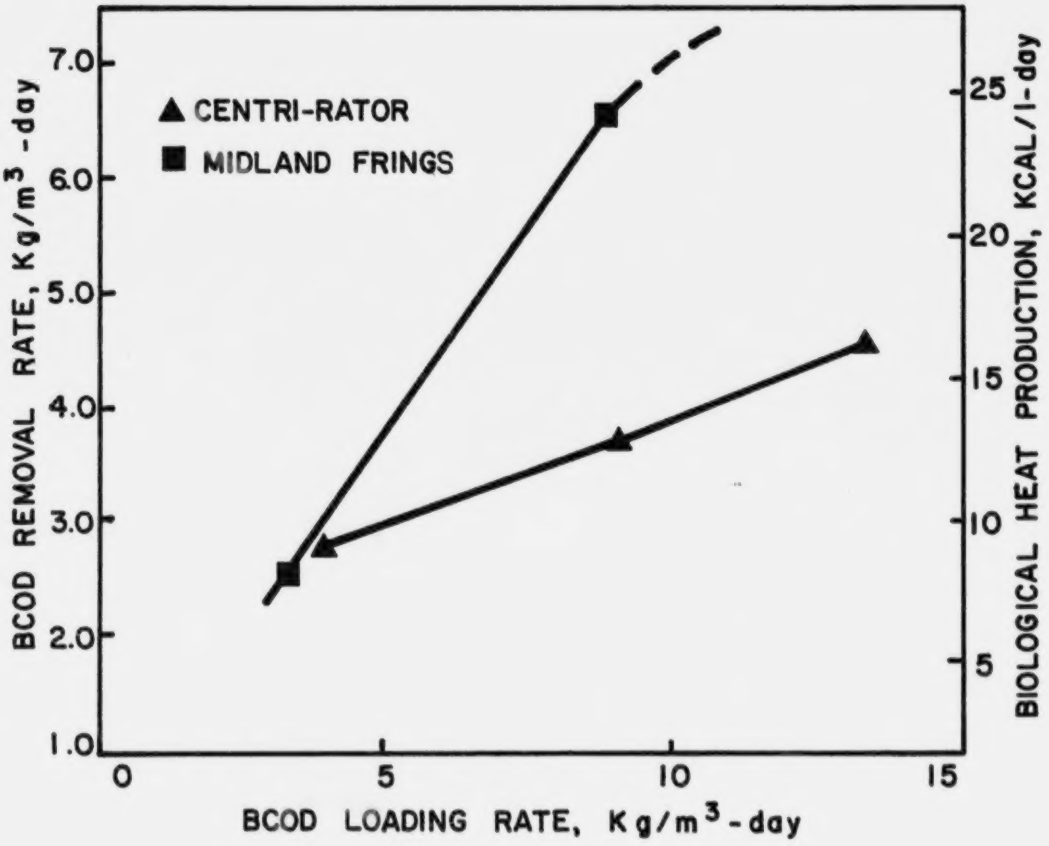


Figure 61. Illustration of organic removal rate (BCOD) and total biological heat production (BHP) as related to organic loading rate for the full scale system at steady state conditions.

Figure 62. The relationship of R to loading rate for the Midland-Frings was shown to be the inverse of that observed for the centri-rator. These differences will be discussed later. It is pertinent to note here that R is a critical parameter in determining autoheating potential of a substrate.

### Nitrogen--

The lack of nitrification at high temperatures allowed the conservation of nitrogen through the system at all loadings (see Table 16 for details). The largest nitrogen loss was 28 percent at a 5-day HRT. Typical feed concentrations were 2,000 mg/ℓ as TKN, 100 mg/ℓ as $NH_3$-N with all losses being attributed to ammonia stripping through aeration.

### Loading Rate - pH Relationship--

The resulting pH of high temperature biological reactors is dependent upon the carbon dioxide production rate, ammonia levels, and other factors. Since the aerators were similar in terms of their total oxygen transfer capability, all resulting pH values should be comparable and are shown in Figure 63. The high pH values at the long retention times reflect higher ammonia values, lower carbon dioxide production rates, and more efficient stripping of dissolved carbon dioxide. The net result is a higher pH at low loading rates and long hydraulic retention times, and a lower pH for the opposite conditions. The lower pH values also assist in explaining the low ammonia losses that occurred at loading rates greater than 10 kg/m$^3$-day.

### Dewaterability and Settleability--

Table 16 illustrates that dewaterability as measured by capillary suction time (CST) deteriorated at all conditions studied. The largest decrease in this dewaterability index occurred at high loading rates and probably was a function of the microbial population and high solids concentration.

### Dissolved Oxygen Levels--

During initial operation with the centri-rator aerator, the reactor tank did not contain a curved side at the bottom called a radius recommended by the manufacturer, and the unit was batch fed on a once daily basis. It was thought that the internal mixing patterns would not be adversely affected by the lack of this smoothing of the tank corners or the feeding method. Routine dissolved oxygen analyses of the reactor showed significant concentrations in the top two to three feet with essentially zero values below this depth. Initially, temperature profile also reflected this same trend, i.e., higher temperatures occurred above this level, with decreasing values below three feet. Clearly, this indicated that the floating centri-rator had limited capability for mixing the entire reactor, resulting in better mixing in the top layers with high D.O. values and stagnant lower layers with low D.O. values. This aerator had been shown to have oxygen transfer capability higher than it was demonstrating (6,7).

Operation of the reactor without the radius and with batch daily feed at a 5-day HRT is shown in Figure 64. A decreased loading rate that resulted from a 7-day HRT increased the top layer D.O., but the unit was still not exhibiting completely mixed characteristics.

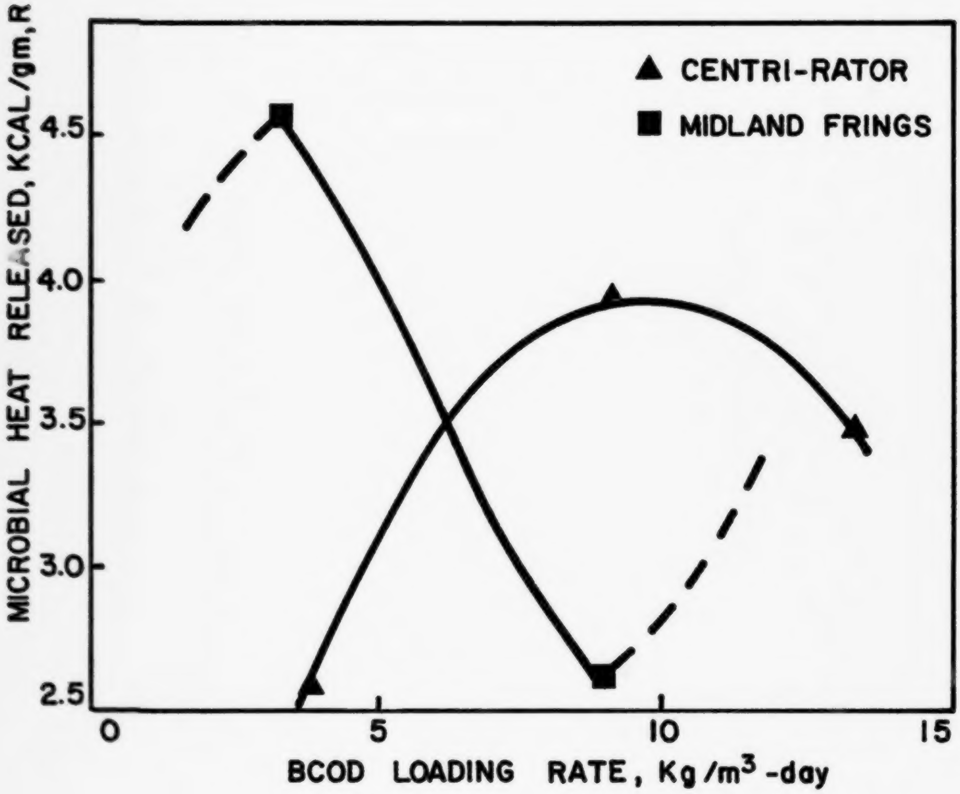


Figure 62. Quantity of heat released during microbial respiration as a function of BCOD loading for both the centri-rator and Midland-Frings aerators.

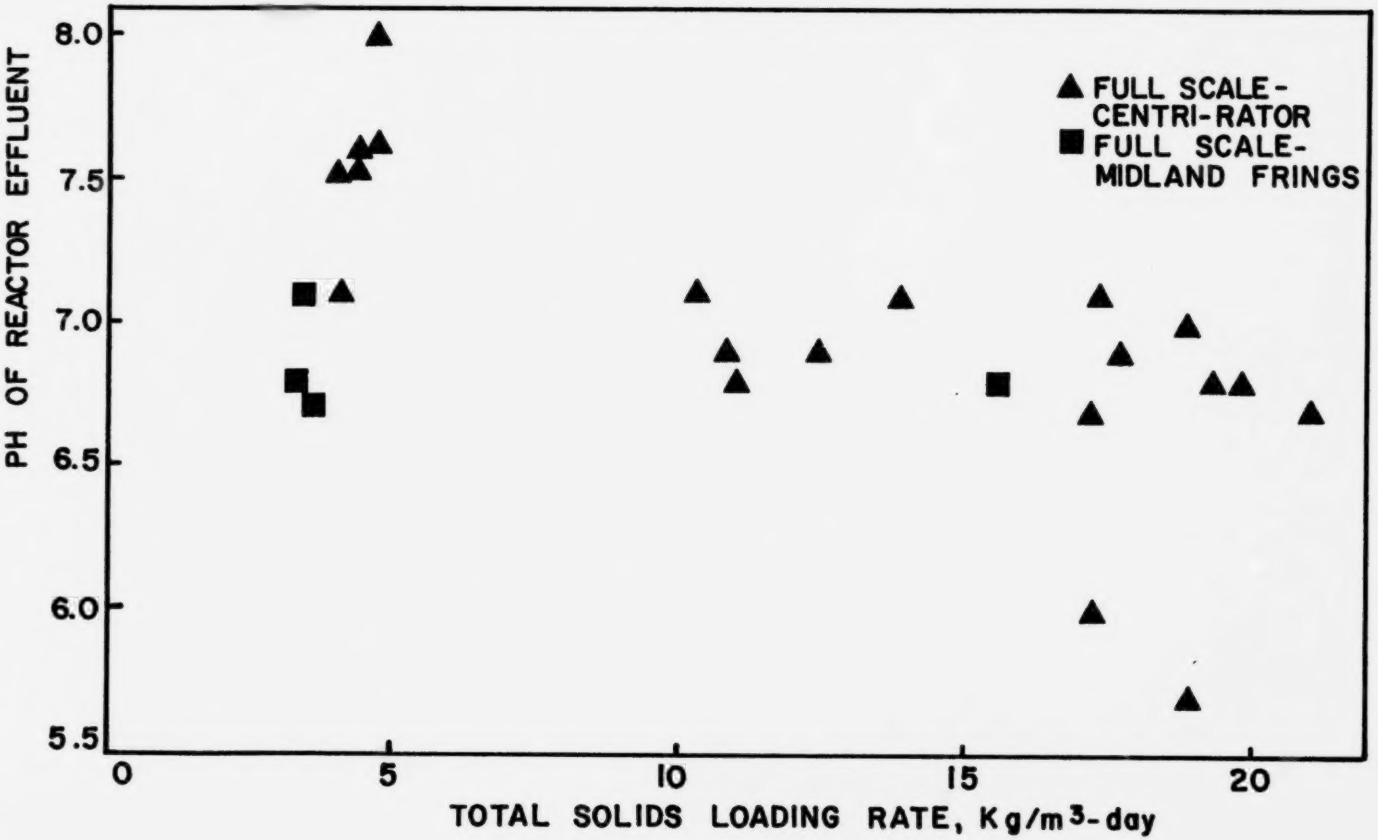


Figure 63. Observed relationship between pH and solids loading rates for full scale study.

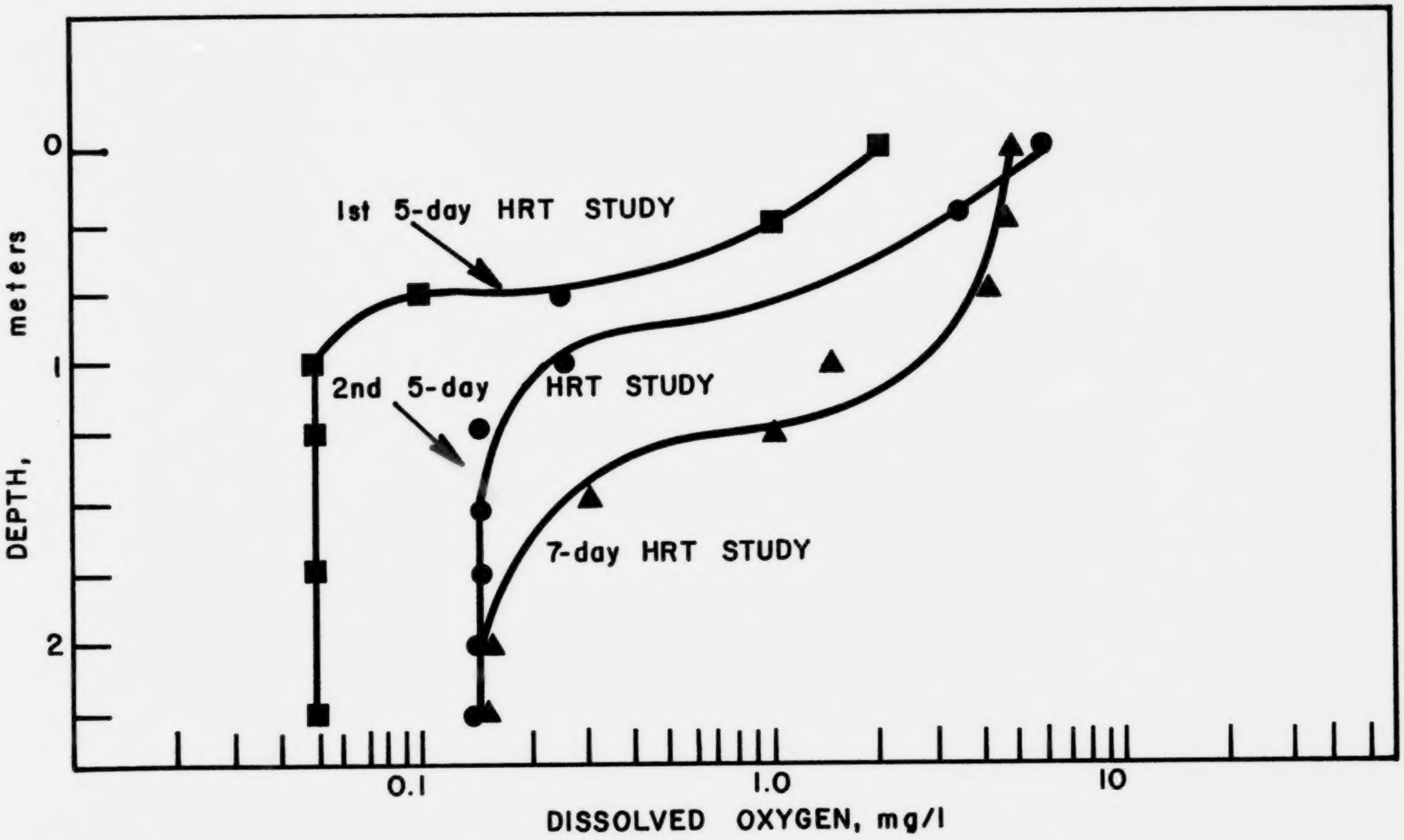


Figure 64. Distribution profiles of dissolved oxygen concentrations with depth within the reactor at various organic loading rates.

At the beginning of the second year of the project, concrete was added to the reactor to fill the corners to attempt to improve the mixing characteristics. The data shown in Figure 65 are the average results that were obtained with rounded reactor corners for both aerators. The data for dissolved oxygen measurements with depth for all full scale data points are summarized in Appendix E. The continued limitations of D.O., resulting from the relatively low rate of mixing, was reflected in the operational results after the unit was modified.

By observation, the Midland-Frings aerator appeared to impart much greater mixing to the slurry than the centri-rator. It should also be noted that the Midland-Frings was located close to the bottom of the reactor. The data shown in Figure 65 indicate that the reactor was operating at higher D.O. values with the Midland-Frings unit. However, direct comparison of the two aerators was not possible because of the lower loading rates studied with the Midland-Frings unit.

Aerator Performance--
The characteristics of the two aerators tested were evaluated in terms of transfer efficiency and transfer rate in sewage sludge under steady state, continuous operation. The majority of testing was conducted with the centri-rator with the Midland-Frings being available during the last seven months of the study. The results are summarized in Table 18. Both gas chromatograph and TCOD removal rate methods (as described in Section 5) were used to determine the transfer efficiency and rate of each aerator. The results of these two methods correlated well throughout the study. The highest transfer efficiency was observed at the highest loading rates for both aerators. This transfer efficiency was 22 percent at a loading rate of 9 kg BCOD/$m^3$-day for the Midland-Frings and 19 percent at a loading rate of 13 kg BCOD/$m^3$-day for the centri-rator, as shown in Figure 66. The oxygen transfer rate exhibited a similar trend with 1.8 kg $O_2$/kW-hr (2.7 hp $O_2$/kW-hr) as the maximum for the Midland-Frings and 1.7 kg $O_2$/kW-hr for the centri-rator (Figure 67). The centri-rator exhibited a sevenfold increase in oxygen transfer efficiency over its tap water performance. This clearly indicated the lack of usefulness of the tap water tests for evaluation of aerator potential.

The airflow through the centri-rator was found to vary with the loading rate as shown in Figure 68. The average airflow during operation with sewage sludge was only 70 percent of the values recorded in tap water. The Midland-Frings showed no variation in airflow with loading, i.e., was constant at 40 ℓ/sec.

## Factors Responsible for Oxygen Transfer Efficiency

As noted in the Introduction, the definition of factors which caused the self-aspirating aerators to achieve oxygen transfer efficiencies more than an order of magnitude greater than would be expected using all other conventional aerators was not a part of the objectives of this study. An earlier study had clearly indicated that the self-aspirating aerators could achieve operational results that differed greatly from what would be predicted using conventional analysis with tap water characterization. It was, however, necessary to attempt to indicate which aeration parameters were

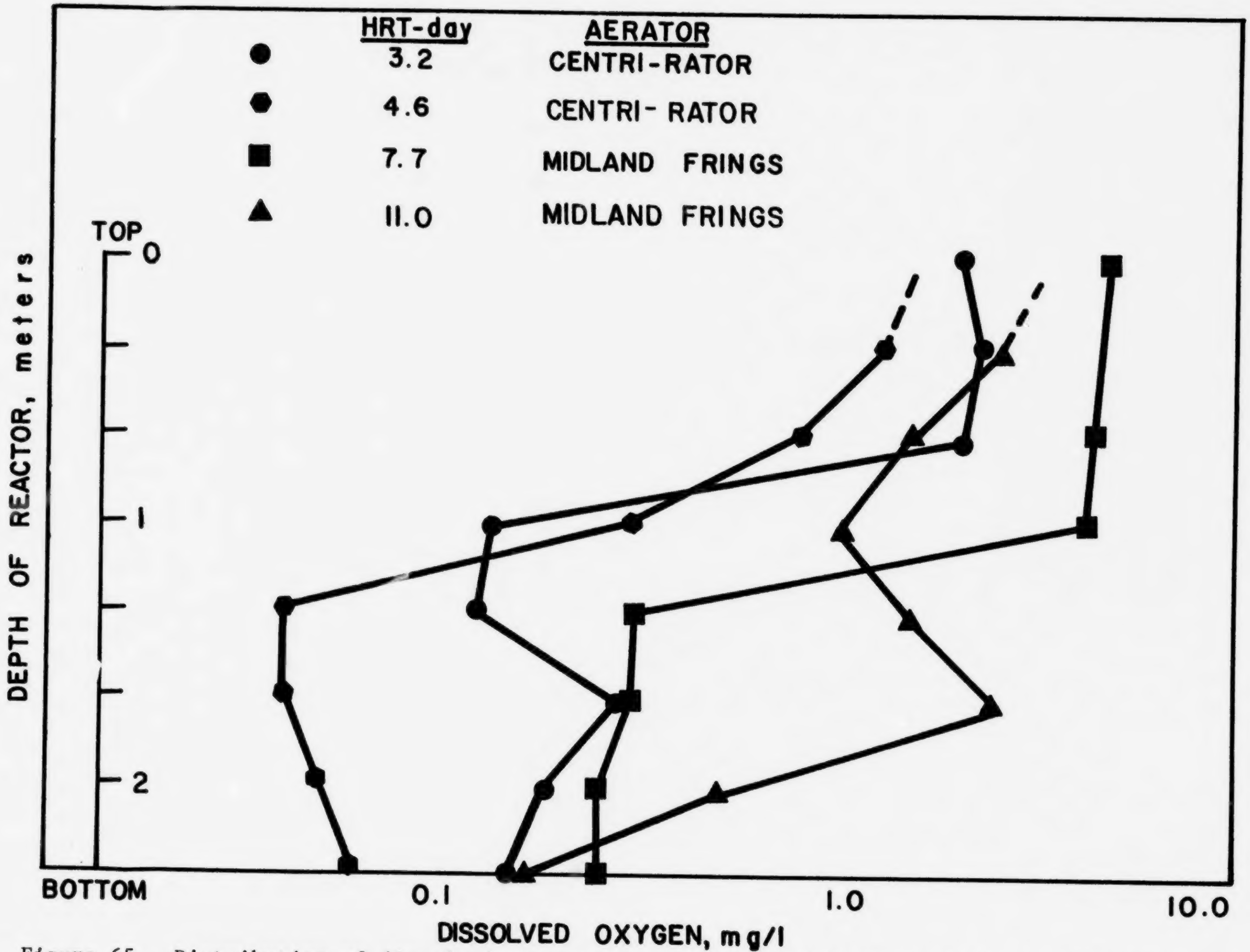


Figure 65. Distribution of dissolved oxygen concentrations with depth within the reactor after the addition of a concrete radius.

## TABLE 18. SUMMARY OF OXYGEN TRANSFER RESULTS FOR STEADY STATE FULL SCALE OPERATION

| | Identification* | | | | |
|---|---|---|---|---|---|
| | C<br>(3.20) | M<br>(3.49) | C<br>(4.56) | C<br>(7.74) | M<br>(11.0) |
| BCOD Loading Rate $kg/m^3$-day | 13.3 | 8.83 | 9.08 | 3.89 | 3.27 |
| Airflow C/sec | 26.0 | 39.6 | 25.5 | 25.2 | 40.0 |
| Oxygen Transfer Efficiency† % | 22.9 | - - | - - | 8.67 | - - |
| Oxygen Transfer Rate† kg/kW-hr | 1.94 | - - | - - | 0.59 | - - |
| Oxygen Transfer Efficiency COD Basis % | 18.9 | 22.1 | 18.3 | 13.8 | 7.90 |
| Oxygen Transfer Rate COD Basis kg/kW-hr | 1.67 | 1.80 | 1.22 | 0.87 | 0.66 |

*C = centri-rator; M = Midland-Frings; ( ) = HRT-day.

†G.C. method.

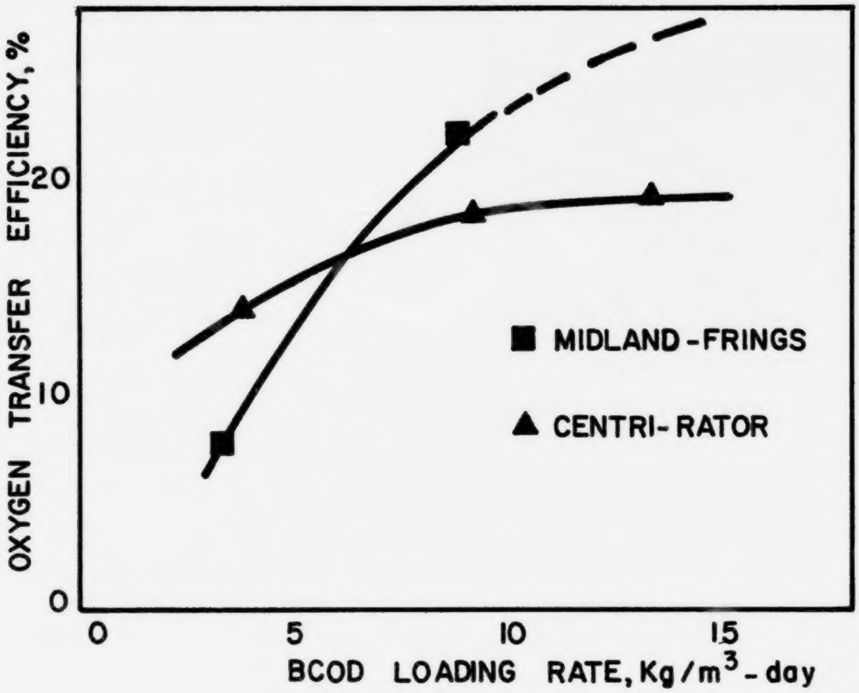


Figure 66. Oxygen transfer efficiency, as measured by TCOD removal-rate analyses, as related to organic loading rate.

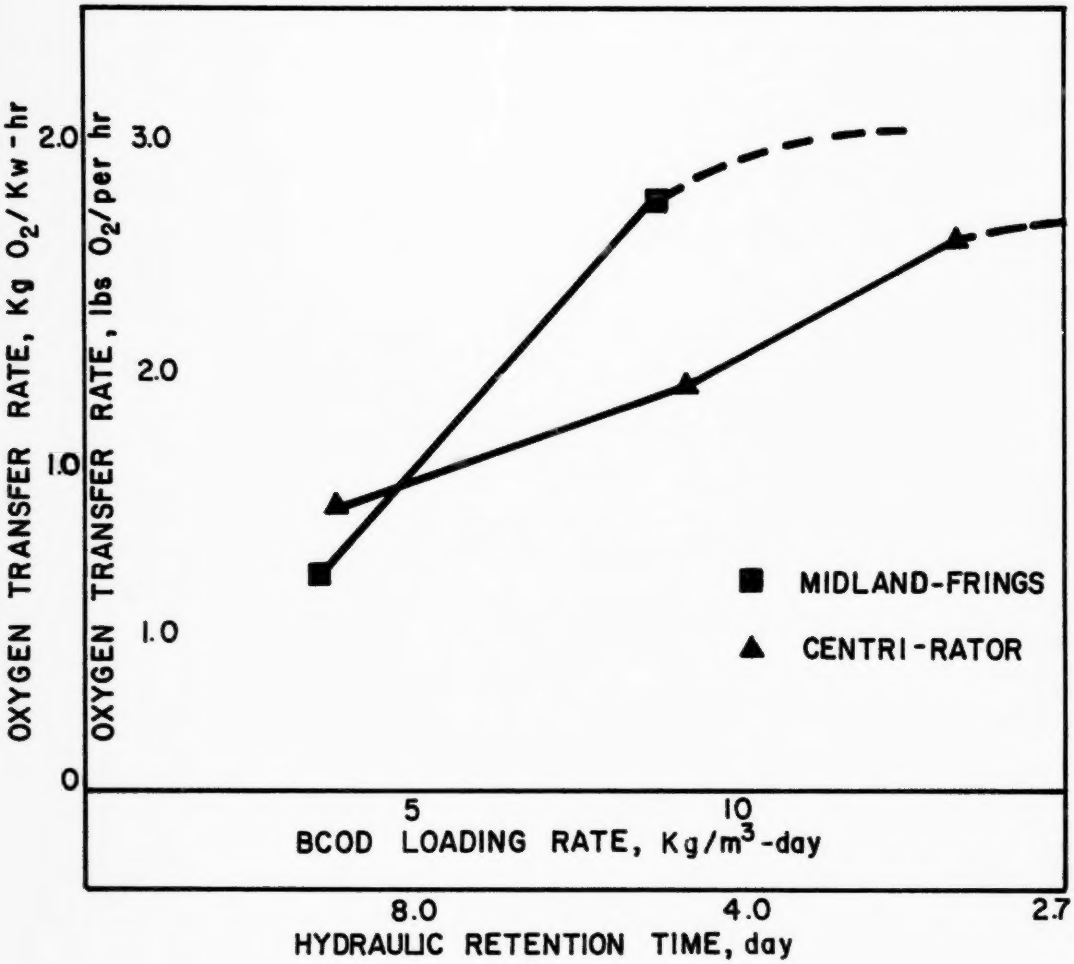


Figure 67. Oxygen transfer rate (O.T.R.) as related to organic loading rate (BCOD) and hydraulic retention time. (HRT based on constant feed concentration of 40 gm BCOD/ℓ.)

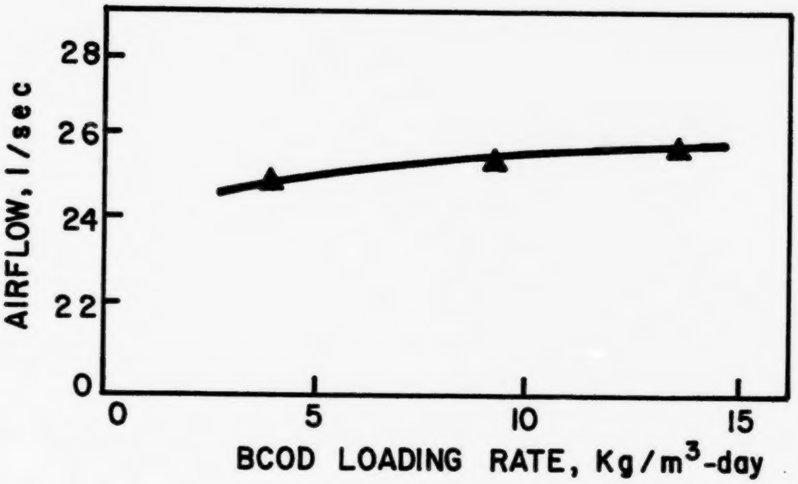


Figure 68. Airflow as related to organic loading (BCOD) rate for the centri-rator.

being affected in order to support the unique nature of the aeration results that are reported here. In order to do this, aeration theory was examined briefly to clarify the controlling parameter(s).

The fundamental aeration relationship is often described by:

$$\frac{dc}{dt} = K_d A \ (C_s - C) \tag{37}$$

where $dc/dt$ = change in oxygen concentration (rate of gas absorption), mass/time

$K_d$ = coefficient of diffusion, time$^{-1}$

$A$ = cross-sectional area through which the gas is transferred, length$^2$

$C_s$ = saturation concentration of the gas in the aerating liquid, mass/volume

$C$ = concentration of the gas at time t, mass/volume.

It is well known that gas transfer is sensitive to temperature changes, and that increased temperatures decrease the effectiveness of oxygen transfer. Equation (37) reflects the reasons why this is true. The absorption coefficient, $K_d$, for air in water decreases to one-half its 0°C value at 30°C. Also, the saturation value for oxygen in water (when aerating with air containing 20.9 percent by volume of oxygen under absolute pressure of 760 mm of mercury) decreases from 14.6 mg/ℓ $O_2$ at 0°C to less than 6 mg/ℓ $O_2$ at thermophilic temperatures around 60°C. Finally, equation (37) is traditionally altered to account for adverse effects of chemicals, solids, and other impurities that tend to decrease the effectiveness of oxygen transfer. Thus, all these factors indicate that aeration of organic slurries at thermophilic condition will be much more difficult than comparable aeration at lower temperatures.

Conversely, factors which are controllable and should enhance the transfer rate are the cross-sectional area, A, of transfer and variation of the saturation value of oxygen, $C_s$. The value of $C_s$ is directly proportional to the partial pressure of oxygen in the aerating gas. Thus, the use of pure oxygen should increase the rate of transfer of oxygen by almost a factor of five over air which is only 20.9 percent oxygen. This is the reason that the use of pure oxygen has been recommended for use in the thermophilic aerobic digestion process. Oxygen transfer is more sensitive to changes in the bubble size than $C_s$, as shown in Figure 69. A decrease in the bubble diameter by a factor of 10 causes the total surface area through which oxygen can be transferred to increase by a factor of 10 for every volume of gas passed through the liquid.

Direct observation of the aerator in tap water and organic slurries indicated that the bubbles entrained in the organic slurries were much smaller than those produced in tap water. In fact, the bubbles were so small that they remained entrained in samples taken from the reactor for several hours after sampling.

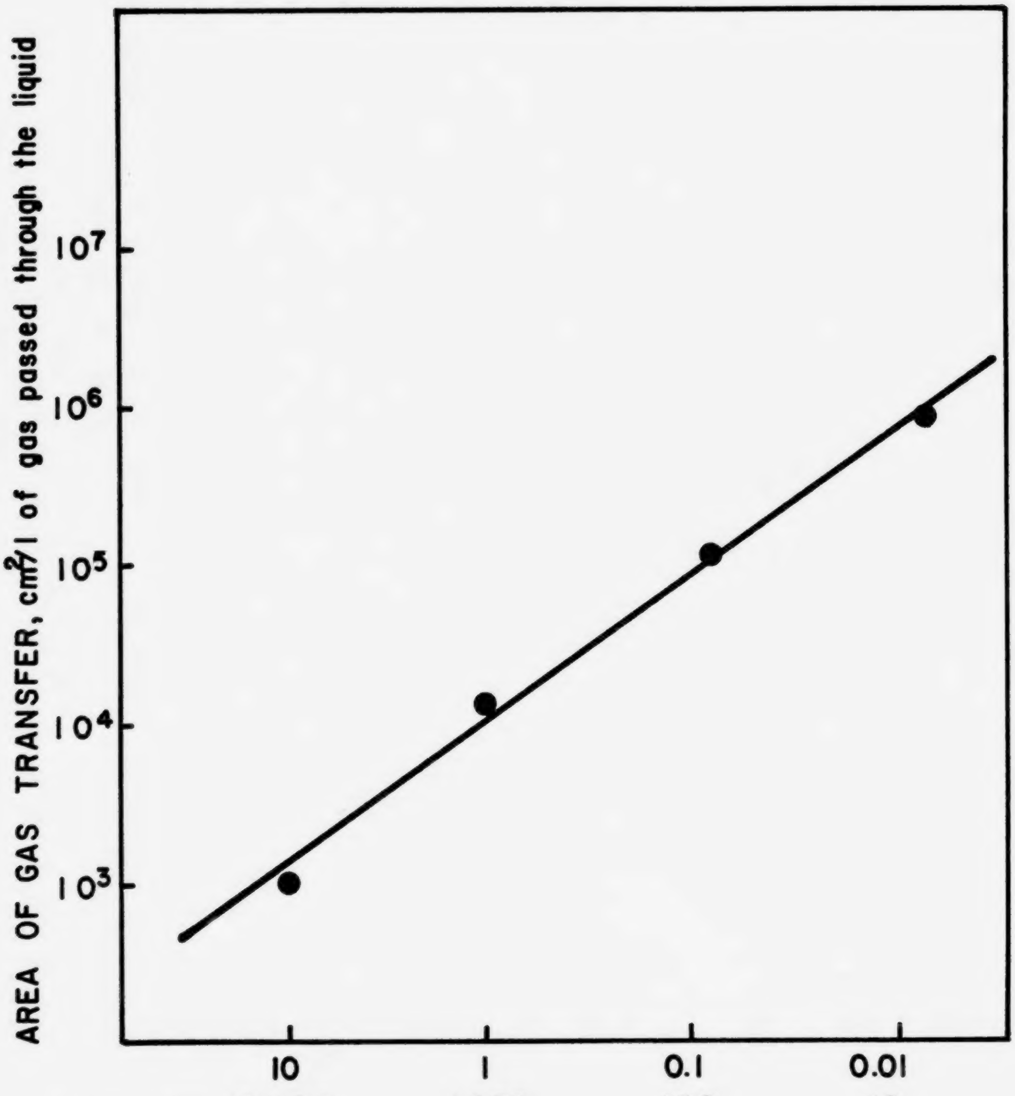


Figure 69. Relationship of bubble diameter to the area available for gas transfer.

In order to document the size of the bubble created by the self-aspirating aerators, the following method was developed. A sampling port had been located on the side of the reactor, 50 cm from the bottom of the tank. Although samples taken from this port may have had bubbles larger than those newly formed near the aerator, it was assumed that those contained in the slurry were representative of most bubbles in the tank. A flat chamber about 900 $cm^2$ by 2.5 cm deep was attached to this port. The slurry was introduced into the chamber by opening the valve and allowing several liters to pass through the chamber. After closing the valve to stop the flow, a photograph of the top surface of the chamber was made. The apparatus used for this procedure is shown in Figure 70.

The bubbles developed in tap water by the centri-rator appeared to approach one cm in diameter. Typical results of bubble analysis shown in Figure 71 indicate the extremely small bubble created by both the centri-rator and the Midland-Frings self-aspirating aerators. Most bubbles had diameters varying from 0.01 cm to 0.05 cm. It should also be noted that these bubbles were so small that some were probably retained in the liquid by horizontal and downward mixing forces that exceeded the upward velocities (buoyancy) of the bubble.

One of the characteristics of the treatment system was the foam that was generated. Figure 72 shows a photograph of the bubbles that composed the foam layer. The mechanical foam cutters controlled the level of the foam layer at a 100 cm thickness. This fine textured foam layer probably had an impact on the thermal characteristics that could not be described with tap water testing.

## PATHOGEN SURVIVAL

One of the primary objectives of this study was to document and compare the fate of various pathogenic organisms in the full scale autoheated aerobic thermophilic digester and the full scale mesophilic anaerobic digester (as compared to the background levels of the various organisms in the thickened raw primary-waste activated sludge). To meet this objective, fifteen sets of pathogen samples were collected and sent to four laboratories (see Section 6) for the various pathogen analyses. The sampling dates, along with the laboratory identifications are given in Table 19.

The conditions of operation of the digesters on the given pathogen sample dates are given in Table 20. Samples were collected under a wide range of operating conditions for the thermophilic aerobic digester with all reactor samples at or above 40°C and a pH of 7.0. Loading rates for the thermophilic digester ranged from 2.6 kg TVS/$m^3$-day to 13 kg TVS/$m^3$-day, with treatment efficiencies varying between 20 and 42 percent TVS destruction, respectively. Feeding of the reactor was usually discontinued between 12 to 24 hours prior to sampling for pathogens. This delay was incorporated to more closely approximate the operation of a unit that would be used for pathogen reduction. Sample numbers 6, 11, 12, and 13 from the thermophilic aerobic digester were collected under marginally operational conditions. That is, the temperature was lower than steady state operations, or the reactor was D.O. limited.

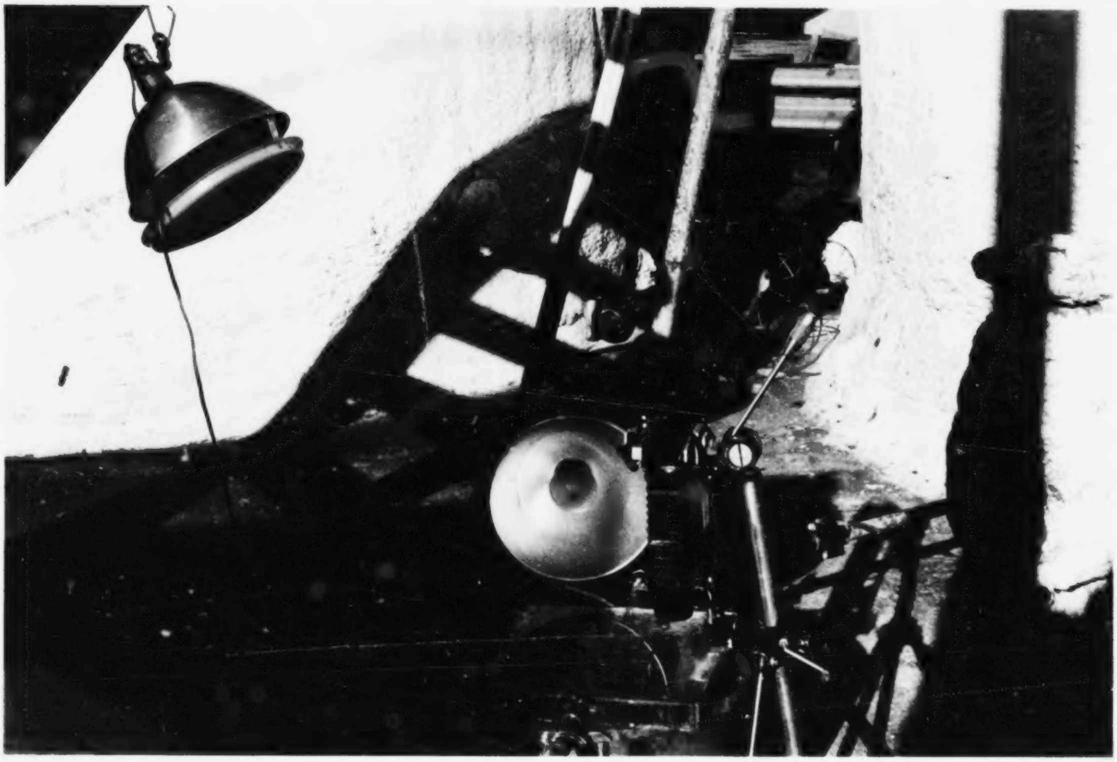


Figure 70: Photograph illustrating the bubble photography apparatus.

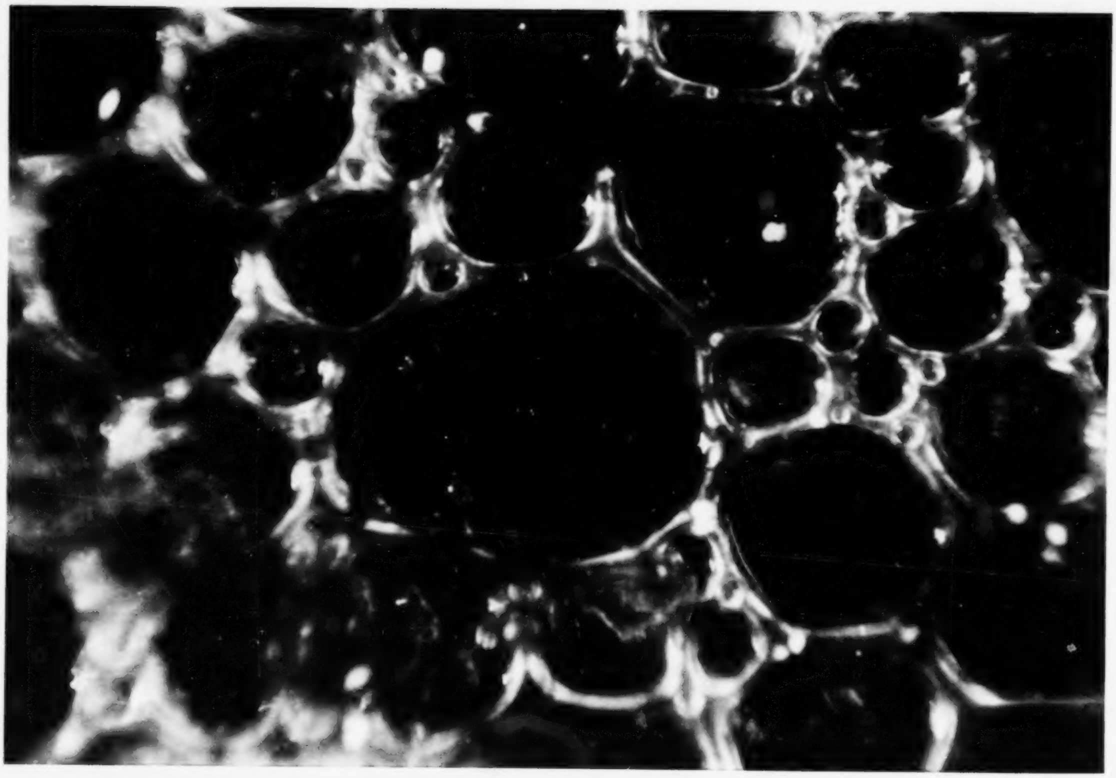


Figure 72: Foam bubbles characteristic of the foam layer generated during autoheated aerobic digestion of sewage sludge.

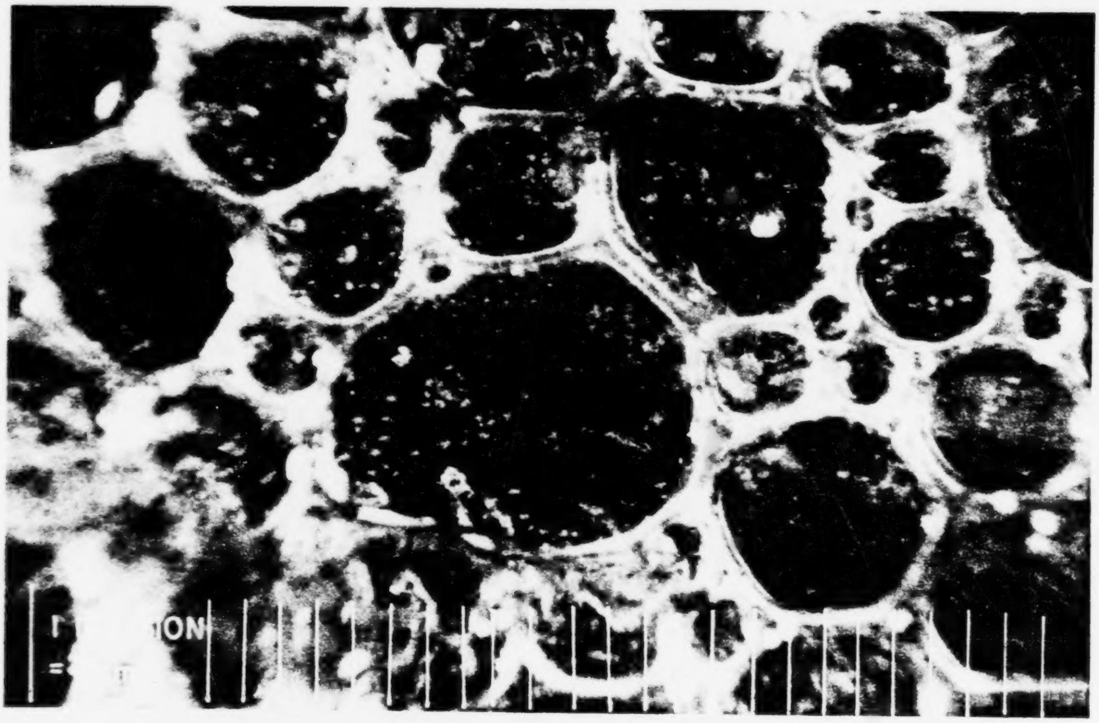


Figure 72: Foam bubbles characteristic of the foam layer generated during autoheated aerobic digestion of sewage sludge.

TABLE 19. PATHOGEN ANALYSES CONDUCTED ON SAMPLES OF RAW PRIMARY AND WASTE ACTIVATED SLUDGE MIXTURES, AND EFFLUENT FROM THE MESOPHILIC ANAEROBIC DIGESTER AND THE THERMOPHILIC AEROBIC DIGESTER

| EPA Laboratory - Cincinnati | | | Cornell University | Tulane University |
|---|---|---|---|---|
| Pathogens (Bacteria) | Viruses | Parasites (Ascaris) | Pathogens (Bacteria) | Parasites |
| 8/15/77 | 8/15/77 | 8/15/77 | * | * |
| 8/29/77 | 8/29/77 | 8/29/77 | * | * |
| 9/26/77 | 9/26/77 | + | * | * |
| + | + | 11/23/77 | * | * |
| + | 3/15/78 | + | * | * |
| 3/27/78 | + | 3/27/78 | * | * |
| + | + | + | * | 4/04/78 |
| + | 4/25/78 | + | * | 4/25/78 |
| + | + | + | * | 5/08/78 |
| + | 5/15/78 | 5/15/78 | * | * |
| + | 5/22/78 | 5/22/78 | 5/22/78 | 5/22/78 |
| 6/26/78 | 6/26/78 | 6/26/78 | 6/26/78 | 6/26/78 |
| + | 8/07/78 | 8/07/78 | 8/07/78 | 8/07/78 |
| + | 8/14/78 | 8/14/78 | * | 8/14/78 |
| 11/13/78 | 11/13/78 | 11/13/78 | * | * |

*No samples taken or shipped.

+No analysis due to one of the following:
(1) breakage during shipment
(2) delayed shipment
(3) excessive temperature of sample.

## TABLE 20. OPERATIONAL DATA FOR DATES OF PATHOGEN SAMPLING.

| | | THERMOPHILIC AEROBIC DIGESTER | | | | | MESOPHILIC ANAEROBIC DIGESTER | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Sample No. | Date of Sample Collection | HRT days | Volatile Solids Loading Rate kg/$m^3$-day | Volatile Solids Treatment Efficiency % | Reactor Temp. °C | pH Reactor | Feed Temp. °C | pH Feed | HRT days | Gas Production Vol. Gas / Vol. Reactor | pH Reactor | Reactor Temp. °C |
| 1 | 8/15/77 | 8.02 | 2.96 | 42.1 | 48.5 | 7.7 | 22.4 | 5.4 | 19.9 | 0.48 | 6.4 | 35.0 |
| 2 | 8/29/77 | 7.20 | 2.92 | 38.0 | 45.2 | 7.8 | 21.4 | - - | - - | - -[b] | - - | - - |
| 3 | 9/26/77 | 4.88 | 4.41 | 19.5 | 47.1 | 7.9 | 20.0 | 6.0 | - - | - - | - - | - - |
| 4 | 3/15/78 | 4.39 | 6.98 | 28.7 | 48.0 | 7.2 | 6.5 | 5.8 | 19.2 | 0.65 | 6.8 | 35.0 |
| 5 | 3/27/78 | 5.51 | 5.71 | 37.8 | 49.7 | 7.2 | 5.5 | 5.7 | 30.6 | 0.72 | 6.9 | 35.0 |
| 6 | 4/04/78 | ≈5.0[c] | ≈6.0 | ≈30 | 55.0 | - - | 5.0 | - - | 24.7 | 0.52 | 6.9 | 35.0 |
| 7 | 4/25/78 | 3.03 | 13.5 | 28.7 | 54.9 | 7.3 | 9.1 | 5.6 | 37.6 | 0.45 | 7.2 | 35.0 |
| 8 | 5/08/78 | 3.51 | 9.76 | 15.5 | 42.0 | 7.0 | 10.8 | 5.9 | 18.9 | 0.48 | 7.0 | 35.0 |
| 9 | 5/15/78 | 3.64 | 9.22 | 22.6 | 52.9 | 7.2 | 12.1 | 5.6 | 20.2 | 0.48 | 7.1 | 35.0 |
| 10 | 5/22/78 | 3.30 | 10.4 | 18.9 | 45.8 | 7.1 | 13.2 | 5.6 | 19.7 | 0.45 | 6.9 | 35.0 |
| 11 | 6/26/78 | 3.60 | d | d | 40.0 | d | 20.0 | d | 23.0 | 0.34 | 7.1 | 35.0 |
| 12 | 8/07/78 | ≈4[e] | ≈8.0 | ≈35 | 50.0 | - - | 20.0 | - - | 30.6 | 0.28 | 7.1 | 35.0 |
| 13 | 8/14/78 | ≈4[e] | ≈8.0 | ≈35 | 55.0 | - - | 21.0 | - - | 30.6 | 0.28 | 7.1 | 35.0 |
| 14 | 11/13/78 | 11.0 | 2.60 | 33.2 | 65.0 | 8.3 | 11.0 | 5.6 | 24.3 | 0.45 | 6.9 | 35.0 |

[a] calculations from Binghamton-Johnson City STP records.

[b] no anaerobic sample

[c] feed pump problems

[d] system marginally operational

[e] aerator problems

However, as will be shown later, this did not significantly affect the pathogen inactivation ability of the system.

The mesophilic anaerobic digester was operated under fairly constant conditions with reactor temperature stable at around 35°C, and reactor pH varying between 6.4 and 7.1 (see Section 5 for a detailed summary of anaerobic digester performance).

## Pathogenic Bacteria

Results of the analyses for bacteria are presented in Table 21, with a summary of the results of the tests for fecal coliforms and fecal streptococci shown in Table 22. Historically, fecal coliforms and fecal streptococci have been used as indicator organisms for water quality. However, in this study the fate of these organisms was of secondary interest with the primary focus on the fate of *Salmonella* sp. and *Pseudomonas aeruginosa*. For all dates of analyses, the number of fecal coliforms and fecal streptococci were reduced at least an order of magnitude more than the levels in the raw sludge for both the anaerobic and aerobic digesters. The thermophilic aerobic digester demonstrated even larger reductions, one to three orders of magnitude, of these organisms over the anaerobic digester.

The number of *Salmonella* sp. found in all samples analyzed is given in Table 21, and graphically shown in Figure 73. On all of the sample dates, with the exception of 8/15/77, the aerobic digester effluent showed far greater reductions in *Salmonella* sp. than the anaerobic digester effluent. Analyses for *Pseudomonas aeruginosa*, a facilitative human pathogen, are presented in Table 21. *Pseudomonas aeruginosa* is an opportunistic pathogen mostly associated with nosocomial infections, especially in people with suppressed immune systems, therefore of importance to anyone with a compromised immune system that would be exposed to sludge material containing large numbers of pseudomonads. The results of the analysis for pseudomonads are graphically depicted in Figure 74. In most cases the thermophilic aerobic digester exhibited larger reductions in pseudomonads than the conventional mesophilic anaerobic digester.

The Cornell University Veterinary Medicine Diagnostic Laboratory screened sludge samples on three separate occasions for *Enterobacteriaceae* and related enteric pathogens. Only one such organism was isolated, on 6/26/78, from the aerobic digester, and on 8/7/78, from the anaerobic digester. The organism was *Citrobacter freundii*, as shown in Table 23. This organism is common to intestinal tracts of healthy individuals but can act as an opportunistic pathogen in mass alimentary infections, and in infections of the meninges, middle ear, gall bladder, and the urinary tract. More will be said about the low number of pathogens found in these screenings in a subsequent section.

## Viruses-Total Plaque Forming Units

The results of virus survival tests or total plaque forming units (PFU) per 100 ml wet sludge for raw, anaerobic, and aerobic samples on all dates of collection are shown in Table 24. A histogram representing total PFUs in all

TABLE 21. RESULTS OF BACTERIAL ANALYSES FOR ALL TEST DATES. ALL ANALYSES CONDUCTED BY THE U.S. ENVIRONMENTAL PROTECTION AGENCY LABORATORY, CINCINNATI, OHIO

| | | Count per 100 ml Wet Sludge Sample | | | | |
|---|---|---|---|---|---|---|
| | | Pathogens | | | | |
| Date of Collection | Sludge Type | Salmonella sp. | Pseudomonas aeruginosa | Fecal Coliforms (MF) | Total Coliforms (MF) | Fecal Streptococci |
| 8/15/77 | Raw (22)* | 360 | $1.2 \times 10^{3}$ | $1.5 \times 10^{7}$ | $6.0 \times 10^{8}$ | $2.2 \times 10^{7}$ |
| | Anaerobic (35) | 3 | 3.6 | $1.1 \times 10^{4}$ | $1.3 \times 10^{5}$ | $1.1 \times 10^{4}$ |
| | Aerobic (49) | <3 | 240 | $<3.3 \times 10^{3}$ | $1.7 \times 10^{4}$ | $3.3 \times 10^{3}$ |
| 8/29/77 | Raw (21) | 23 | 700 | $1.0 \times 10^{7}$ | $5.0 \times 10^{6}$ | $3.3 \times 10^{5}$ |
| | Anaerobic (35) | N.S. | N.S. | N.S. | N.S | N.S |
| | Aerobic (45) | <3 | 150 | $5.0 \times 10^{3}$ | $1.4 \times 10^{4}$ | $7.5 \times 10^{3}$ |
| 9/26/77 | Raw (20) | 640 | 440 | $2.0 \times 10^{7}$ | $1.8 \times 10^{9}$ | $3.0 \times 10^{7}$ |
| | Anaerobic (35) | 93 | 72 | $7.6 \times 10^{5}$ | $3.7 \times 10^{7}$ | $9.3 \times 10^{5}$ |
| | Aerobic (47) | <3 | <3 | $<1.0 \times 10^{3}$ | $2.5 \times 10^{4}$ | $2.3 \times 10^{3}$ |
| 3/27/78 | Raw (6) | 2,200 | 16 | $4.4 \times 10^{7}$ | $1.2 \times 10^{9}$ | $2.4 \times 10^{7}$ |
| | Anaerobic (35) | 1,100 | 15 | $2.0 \times 10^{6}$ | $3.5 \times 10^{7}$ | $2.2 \times 10^{6}$ |
| | Aerobic (50) | <3 | <3 | $2.7 \times 10^{4}$ | $3.3 \times 10^{4}$ | $2.3 \times 10^{3}$ |
| 6/26/78 | Raw (20) | 41 | 16 | $3.7 \times 10^{7}$ | $8.7 \times 10^{8}$ | $4.0 \times 10^{7}$ |
| | Anaerobic (35) | 23 | 15 | $3.5 \times 10^{5}$ | $2.0 \times 10^{6}$ | $2.0 \times 10^{5}$ |
| | Aerobic (40) | <3 | <3 | $7.0 \times 10^{4}$ | $1.0 \times 10^{5}$ | $1.8 \times 10^{5}$ |
| 11/13/78 | Raw (11) | 430 | 11 | $9.6 \times 10^{7}$ | $6.0 \times 10^{8}$ | $5.0 \times 10^{7}$ |
| | Anaerobic (35) | 36 | <3 | $1.1 \times 10^{6}$ | $6.4 \times 10^{6}$ | $1.3 \times 10^{6}$ |
| | Aerobic (65) | <3 | <3 | <1,000 | <1,000 | <1,000 |

*( ) = temperature of sample when collected (°C).

TABLE 22. SUMMARY OF THE FATE OF FECAL COLIFORMS AND FECAL STREPTOCOCCI DURING MESOPHILIC ANAEROBIC DIGESTION AND THERMOPHILIC AEROBIC DIGESTION OF PRIMARY AND WASTE ACTIVATED SLUDGE MIXTURES

| | Counts of Organisms per 100 ml Wet Sludge | | | |
|---|---|---|---|---|
| | Fecal Coliforms | | Fecal Streptococci | |
| Sludge Type | Mean | Range | Mean | Range |
| Feed Sludge | $3.7 \times 10^7$ | $1.0-9.6 \times 10^7$ | $2.7 \times 10^7$ | $3.3 \times 10^5-5.0 \times 10^7$ |
| Anaerobic Sludge | $8.4 \times 10^5$ | $1.1 \times 10^4-2.0 \times 10^6$ | $9.3 \times 10^5$ | $1.1 \times 10^4-2.2 \times 10^6$ |
| Aerobic Sludge | $1.8 \times 10^4$ | $1.0 \times 10^3-7.0 \times 10^4$ | $8.2 \times 10^4$ | $1.0 \times 10^3-3.0 \times 10^6$ |

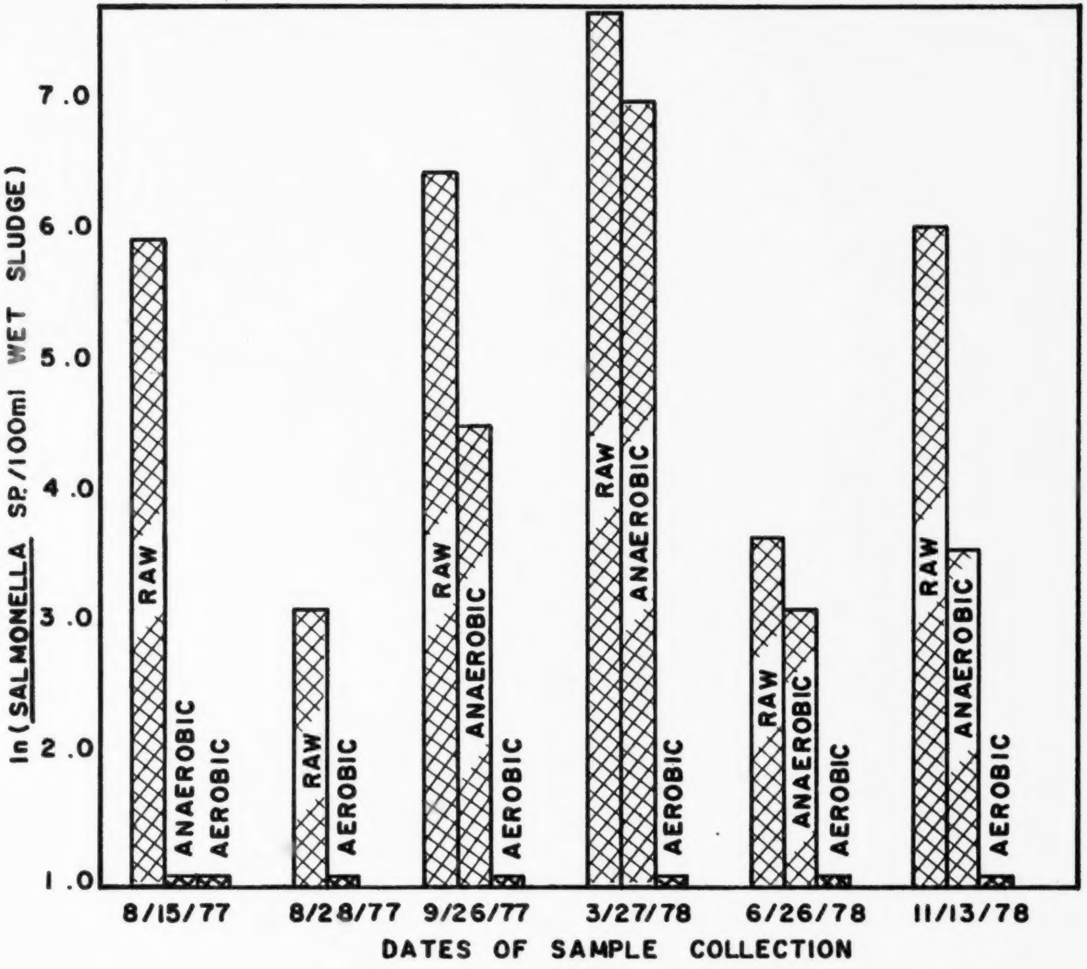


Figure 73. Salmonella sp. as found in raw, anaerobic, and aerobic sludges. ($\leq$3 organisms/100 ml considered below limits of detection.)

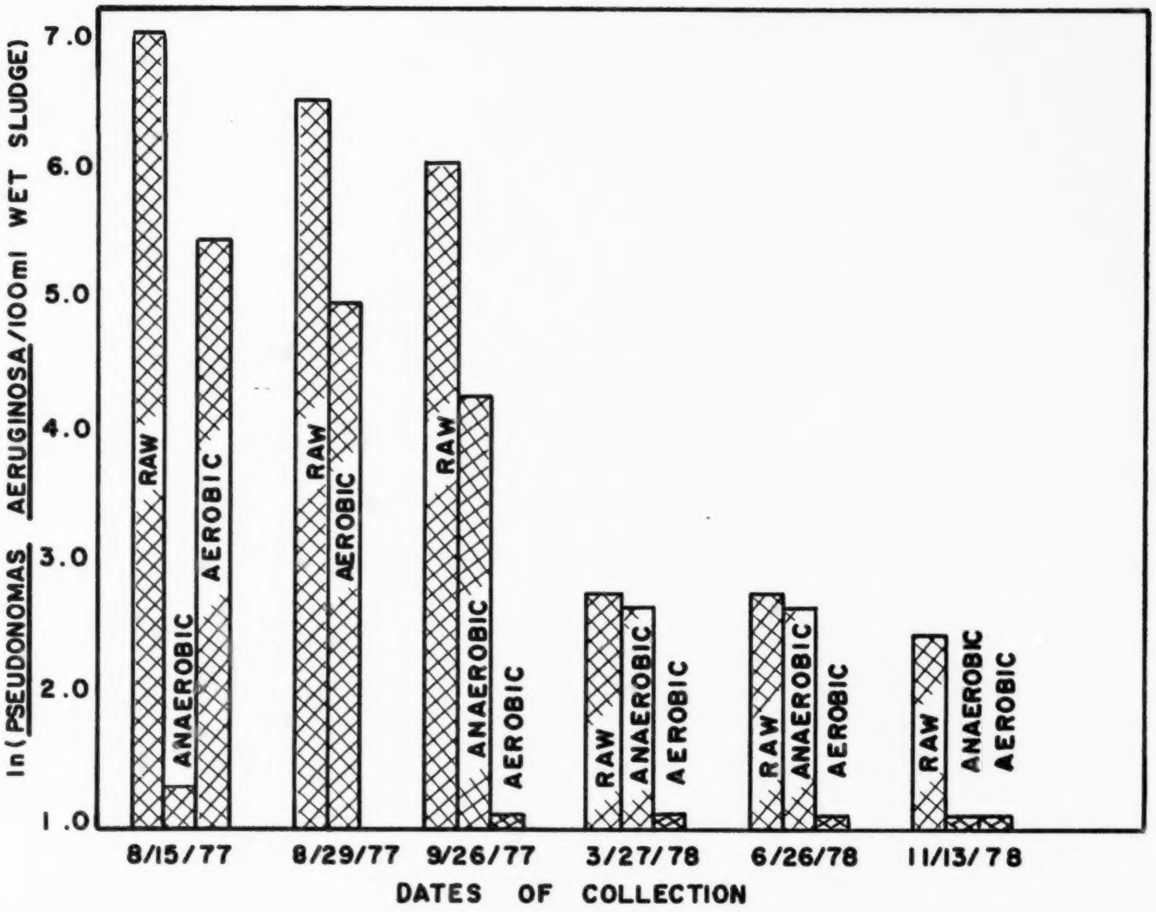


Figure 74. Results of analyses for Pseudomonas aeruginosa in raw, anaerobic and aerobic sludges.

TABLE 23. RESULTS OF QUALITATIVE ANALYSIS FOR ENTERIC PATHOGENS BY CORNELL UNIVERSITY VETERINARY DIAGNOSTIC LABORATORY
(100 ml wet sludge analyzed)

|  | Enteropathogenic Bacteria Isolated |  |  |
|---|---|---|---|
| Date of Collection | Raw Sludge | Anaerobic Sludge | Aerobic Sludge |
| 5/22/78 | None Isolated | None Isolated | None Isolated |
| 6/26/78 | None Isolated | None Isolated | *Citrobacter freundii* |
| 8/07/78 | None Isolated | *Citrobacter freundii* | None Isolated |

## TABLE 24. RESULTS OF VIRUS ANALYSES FOR TOTAL PFU OF PRIMARY AND WASTE ACTIVATED SLUDGE MIXTURES. ANALYSES CONDUCTED BY THE U.S. ENVIRONMENTAL PROTECTION AGENCY LABORATORY

| | Plaque Forming Units (PFU)/100 ml Wet Sludge Sample | | |
|---|---|---|---|
| Dates of Collection | Raw Sludge | Mesophilic Anaerobic Sludge | Thermophilic Aerobic Sludge |
| 8/15/77 | 533<br>(22)[*] | 8<br>(35) | <1<br>(49) |
| 8/29/77 | 141<br>(21) | N.S.[+]<br>(35) | <1<br>(45) |
| 9/26/77 | 1,000<br>(20) | N.S.<br>(35) | <4<br>(47) |
| 3/15/78 | 356<br>(07) | 10<br>(35) | <1<br>(48) |
| 4/25/78 | 356<br>(09) | 5<br>(35) | <3<br>(55) |
| 5/15/78 | 152<br>(12) | 4<br>(35) | <2<br>(53) |
| 5/22/78 | 316<br>(13) | 16<br>(35) | <2<br>(46) |
| 6/26/78 | 462<br>(20) | 50<br>(35) | <2<br>(40) |
| 8/07/78 | 1,060<br>(20) | 100<br>(35) | <3<br>(50) |
| 8/14/78 | 595<br>(21) | 42<br>(35) | 14<br>(55) |
| 11/13/78 | 570<br>(11) | 72<br>(35) | <2<br>(65) |

[*] ( ) = temperature of sample as collected.

[+] N.S. = No Sample.

samples for all dates of collection is shown in Figure 75. In all cases, the aerobic digester showed higher levels of viral inactivation than did the anaerobic digester with viruses below detectable levels in all but one sample of the aerobic unit effluent. The anaerobic digester exhibited levels of virus inactivation equal to or less than 98.6 percent. No effort was made to distinguish between the various kinds of enteroviruses or other viruses. Nor was any effort made to distinguish the distribution of the viruses between the sludge solids and the liquid.

### Parasites

The results of the parasitic ova analysis for Ascaris, as performed by the University of Cincinnati, are presented in Table 25. Due to the lack of ova recovered from all samples, no differences were noted between aerobic and anaerobic digester samples.

A summary of the parasite analyses as performed by Tulane University for all dates of collection is given in Table 26. Due to the low numbers of the various parasitic ova recovered (see Table 27 for additional details) from each sample, trends in the data were difficult to discern. All of the data were summarized as total viable and total nonviable parasite ova for each sample analyzed. These results are shown graphically in Figure 76. In all samples, with the exception of 5/8/78, the aerobic digester exhibited superior reduction in total viable ova than the anaerobic digester. The significance of the eight different species of parasite ova isolated from the various samples will be discussed in a subsequent section. It should be noted that the distinction between viable and nonviable ova was made solely from morphological observations, e.g., disrupted membranes, without subsequent culturing of the suspect ova for confirmation of the morphological diagnosis.

## FATE OF HEAVY METALS

One of the minor objectives of this study was to determine the impact of the application of the high temperature aerobic digestion process on the heavy metal content of the sludge. At the onset, there was concern that the process conditions could significantly affect the heavy metal distribution. This would be of concern in defining the ultimate disposal options available for this treated sludge.

The raw sludge contained significant concentrations of most heavy metals, with cadmium concentration exceeding 100 ppm in the dry sludge. This compares to a maximum value of 10 ppm often suggested as an upper limit for sludge applied to agricultural land.

The concentration of six heavy metals (cadmium, chromium, copper, nickel, lead, and zinc) were tested in the soluble and particulate phase of the raw sludge, anaerobically digested sludge, and aerobically treated sludges. In general, there was no significant difference between the sludges and the majority of all heavy metals was associated with the particulate phase. Further consideration of the fate of the heavy metals undergoing aerobic and anaerobic biological treatments indicated that the metal species would be

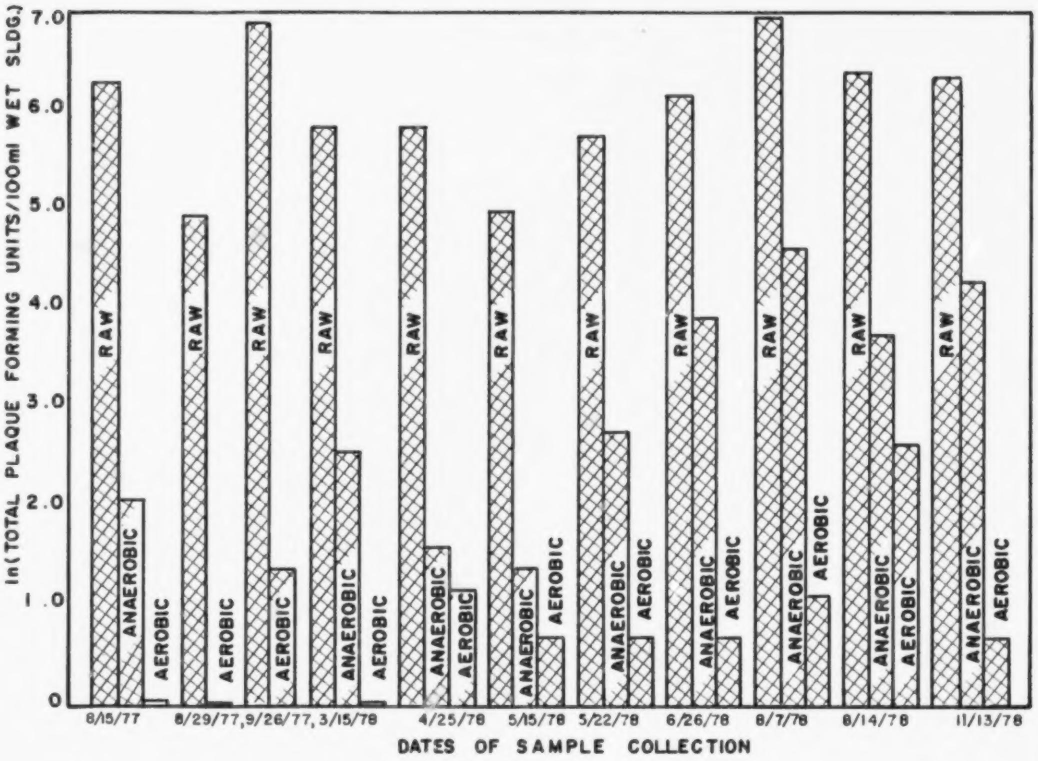


Figure 75. Results of virus analyses (total PFU/100 ml wet sludge) on raw, anaerobically (mesophilic) and aerobically (thermophilic) digested sludge.

TABLE 25. RESULTS OF *Ascaris* ANALYSIS IN PRIMARY AND WASTE ACTIVATED SLUDGE. ALL ANALYSIS CONDUCTED BY THE UNIVERSITY OF CINCINNATI

| | | | *Ascaris* ova/75 gm wet sludge | |
|---|---|---|---|---|
| Dates of Collection | Temperature of Sample When Collected °C | Type of Sludge | Total ova (non-viable and viable) | Viable ova |
| 8/15/77 | 22 | Raw | 0 | 0 |
| | 35 | Anaerobic | 0 | 0 |
| | 49 | Aerobic | 0 | 0 |
| 8/29/77 | 21 | Raw | 0 | 0 |
| | 35 | Anaerobic | N.S.* | N.S.* |
| | 45 | Aerobic | 0 | 0 |
| 11/23/77 | 10 | Raw | 1 | 0 |
| | 35 | Anaerobic | 0 | 0 |
| | 40 | Aerobic | 1 | 0 |
| 3/27/78 | 5.5 | Raw | 1 | 0 |
| | 35 | Anaerobic | 0 | 0 |
| | 50 | Aerobic | 0 | 0 |
| 5/15/78 | 12 | Raw | N.S.* | N.S.* |
| | 35 | Anaerobic | 0 | 0 |
| | 53 | Aerobic | 0 | 0 |
| 5/22/78 | 13 | Raw | 0 | 0 |
| | 35 | Anaerobic | | |
| | 46 | Aerobic | | |
| 6/26/78 | 20 | Raw | 0 | 0 |
| | 35 | Anaerobic | 1 | 0 |
| | 40 | Aerobic | 0 | 0 |
| 8/07/78 | 20 | Raw | | |
| | 35 | Anaerobic | | |
| | 50 | Aerobic | | |
| 8/14/78 | 21 | Raw | | |
| | 35 | Anaerobic | | |
| | 50 | Aerobic | | |
| 11/13/78 | 11 | Raw | 0 | 0 |
| | 35 | Anaerobic | 0 | 0 |
| | 65 | Aerobic | 0 | 0 |

*N.S. = No Sample.

TABLE 26. SUMMARY OF PARASITE ANALYSES FOR ALL DATES OF COLLECTION FOR PRIMARY AND WASTE ACTIVATED SLUDGE COMBINATIONS. ALL ANALYSES CONDUCTED BY THE DEPARTMENT OF TROPICAL MEDICINE, TULANE UNIVERSITY

| Date of Collection | Type of Sludge | Total Parasite Ova Count/100 ml Wet Sludge | | Temp of Sample °C |
|---|---|---|---|---|
| | | Viable | Non-viable | |
| 4/04/78 | Raw | 6 | 2 | 5 |
| | Anaerobic | 18 | 2 | 35 |
| | Aerobic | 1 | 16 | 55 |
| 4/24/78 | Raw | 15 | 6 | 9.1 |
| | Anaerobic | --* | --* | 35 |
| | Aerobic | 0 | 9 | 55 |
| 5/08/78 | Raw | 1 | 8 | 11 |
| | Anaerobic | 3 | 5 | 35 |
| | Aerobic | 4 | 12 | 42 |
| 5/22/78 | Raw | 16 | 1 | 13 |
| | Anaerobic | 9 | 2 | 35 |
| | Aerobic | 1 | 5 | 46 |
| 6/26/78 | Raw | 3 | 4 | 20 |
| | Anaerobic | 5 | 2 | 35 |
| | Aerobic | 1 | 5 | 40 |
| 8/07/78 | Raw | 5 | 1 | 20 |
| | Anaerobic | 7 | 0 | 35 |
| | Aerobic | 0 | 9 | 50 |
| 8/14/78 | Raw | 14 | 0 | 21 |
| | Anaerobic | 4 | 6 | 35 |
| | Aerobic | 2 | 10 | 55 |

*No sample.

TABLE 27. RESULTS OF PARASITOLOGIC EXAMINATIONS OF ALL SAMPLES WITH PRIMARY AND WASTE ACTIVATED SLUDGE COMBINATIONS. ALL ANALYSES CONDUCTED BY TULANE UNIVERSITY

| | | Parasite Counts per 100 ml of Sludge | | | | | | | | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Date of Collection | Sludge Type | *Ascaris* | | *T. trichiura* | | *T. vulpis* | | *Toxocara* | | *H. diminuta* | | *Toxascaris leonia* | | *Capillaria* sp. | | *Entameba coci* (like cysts) | | Temp. of Sample When Collected °C |
| | | V | NV | V | NV | V | NV | V | NV | V | NV | V | NV | V | NV | V | NV | |
| 4/04/78 | Raw | 1 | 0 | 0 | 0 | 1 | 0 | 3 | 2 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | Few | 50 |
| | Aerobic | 0 | 4 | 0 | 1 | 0 | 2 | 1 | 8 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 55 |
| | Anaerobic | 3 | 0 | 2 | 0 | 5 | 0 | 8 | 2 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 35 |
| 4/24/78 | Raw | 4 | 2 | 0 | 1 | 5 | 1 | 5 | 2 | 0 | 0 | 1 | 0 | 0 | 0 | 0 | Few | 9.1 |
| | Aerobic[a] | 0 | 2 | 0 | 0 | 0 | 1 | 0 | 6 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 55 |
| | Anaerobic[b] | - | - | - | - | - | - | - | - | - | - | 0 | 0 | 0 | 0 | 0 | 0 | 35 |
| 5/08/78 | Raw | 0 | 1 | 0 | 0 | 1 | 0 | 0 | 7 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 11 |
| | Aerobic | 1 | 2 | 0 | 1 | 1 | 0 | 2 | 8 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | Few | 42 |
| | Anaerobic | 1 | 1 | 0 | 0 | 1 | 1 | 1 | 3 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 35 |
| 5/22/78 | Raw[c] | 2 | 0 | 0 | 0 | 3 | 0 | 10 | 1 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 0 | 13 |
| | Aerobic | 0 | 3 | 0 | 1 | 0 | 0 | 1 | 1 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 46 |
| | Anaerobic | 3 | 0 | 0 | 0 | 1 | 0 | 5 | 1 | 0 | 0 | 0 | 0 | 0 | 1 | 0 | 0 | 35 |
| 6/26/78 | Raw | 0 | 0 | 0 | 0 | 2 | 0 | 3 | 1 | 0 | 0 | | | | | | | 20 |
| | Aerobic | 0 | 4 | 0 | 0 | 1 | 0 | 0 | 1 | 0 | 0 | | | | | | | 40 |
| | Anaerobic | 0 | 0 | 0 | 0 | 3 | 0 | 2 | 2 | 0 | 0 | | | | | | | 35 |
| 8/07/78 | Raw | 0 | 0 | 0 | 0 | 2 | 0 | 3 | 1 | 0 | 0 | | | | | | | 20 |
| | Aerobic | 0 | 0 | 0 | 4 | 0 | 4 | 0 | 1 | 0 | 0 | | | | | | | 50 |
| | Anaerobic | 0 | 0 | 0 | 0 | 2 | 0 | 5 | 0 | 0 | 0 | | | | | | | 35 |
| 8/14/78 | Raw | 3 | 0 | 1 | 0 | 2 | 0 | 8 | 0 | 0 | 0 | | | | | | | 21 |
| | Aerobic | 2 | 3 | 0 | 0 | 0 | 4 | 0 | 3 | 0 | 0 | | | | | | | 55 |
| | Anaerobic | 0 | 2 | 0 | 0 | 2 | 0 | 2 | 4 | 0 | 0 | | | | | | | 35 |

[a] only 80 ml sludge analyzed.

[b] container broken in transit

[c] only 90 ml sludge analyzed

V = Viable

NV = Non-Viable

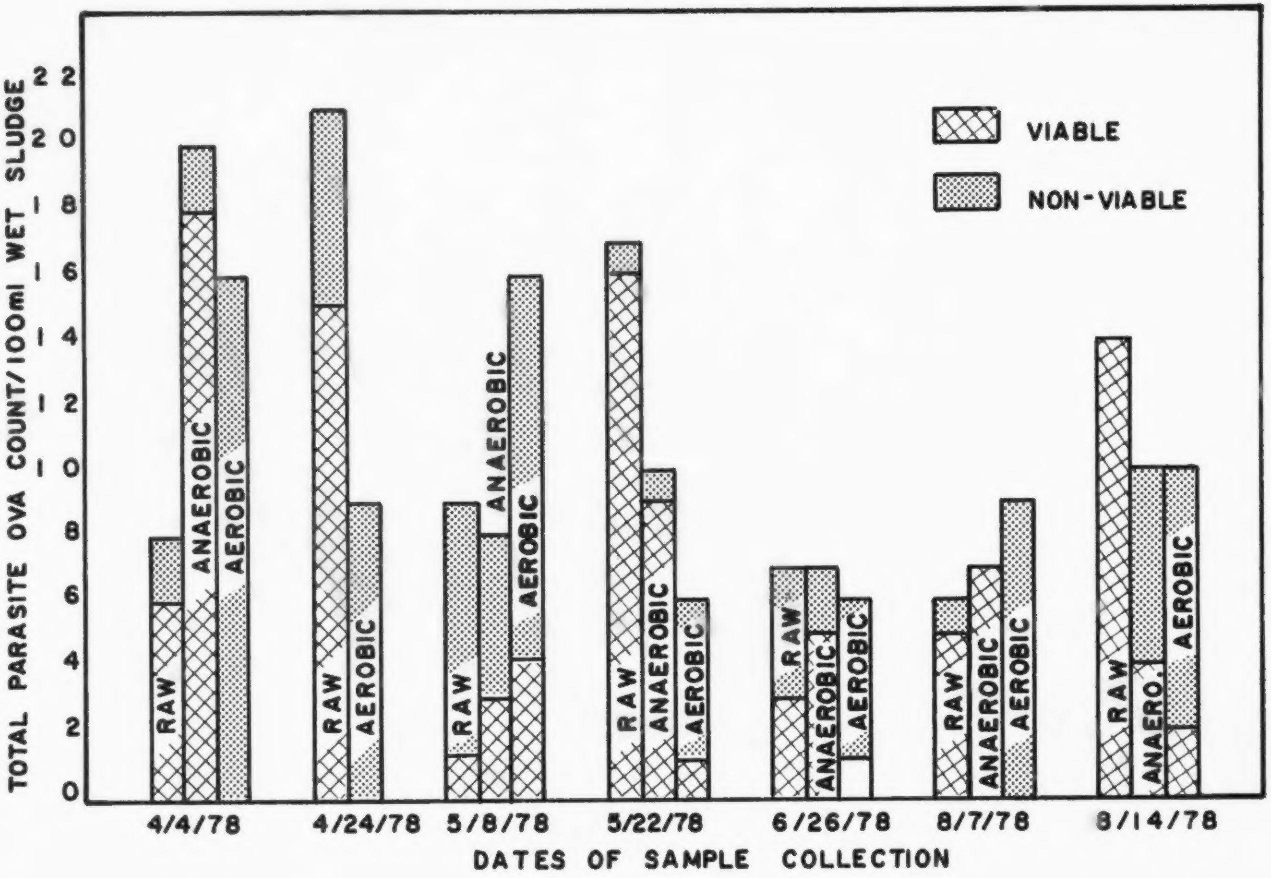


Figure 76. Total Parasite ova found in raw, anaerobic and aerobic sludge for all dates of collection as analyzed by Tulane University.

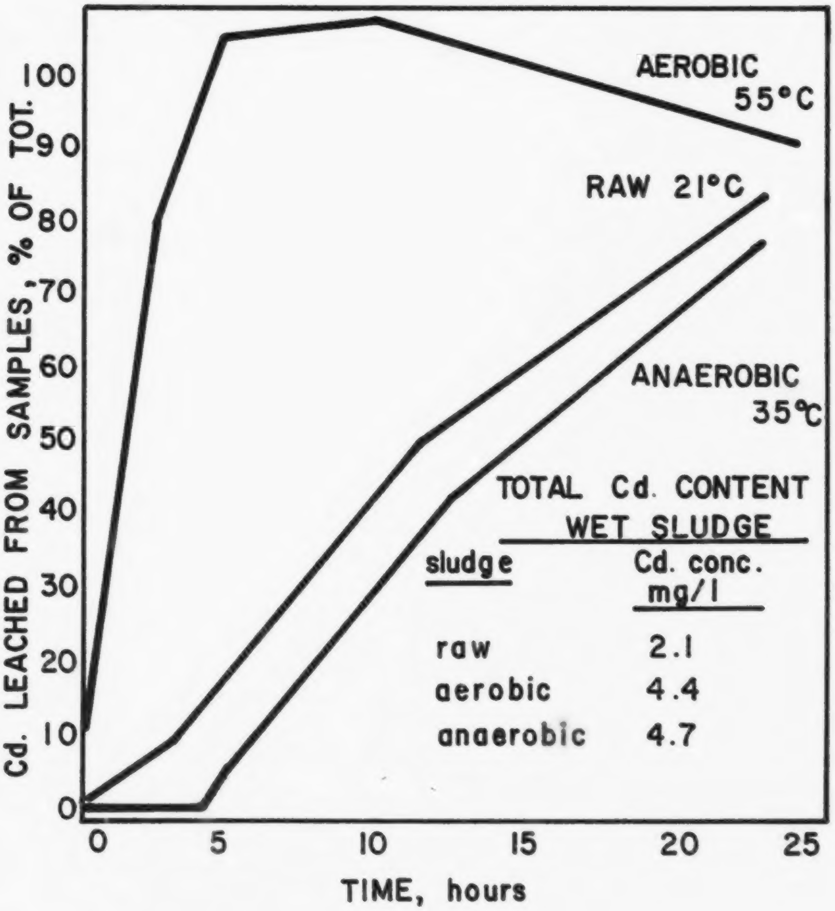
expected to make some adjustment toward a new equilibrium state with the chemical environment established in the reactor with respect to pH, oxidation reduction potential (ORP), the presence of complexing agents, concentrations of precipitant ligands, and temperature. Hayes et al. (10) conducted preliminary comparative studies on the effluents from these systems to determine the potential of removing the heavy metals from the solids. The conversion of heavy metals to the soluble form prior to physical separation would allow the removal of heavy metals with the conventional dewatering techniques in use today (i.e., centrifugation, belt press, dewatering beds, vacuum filter, etc.).

In pure aqueous systems, rapid solubilization of heavy metal precipitates can often be accomplished through acidification. If the solubilization reaction of a heavy metal precipitate is endothermic, higher temperatures will tend to shift equilibrium conditions toward the solubilized species and may enhance the rate of solubilization. The predominant complexes formed and the oxidation state of the heavy metal will also determine the kinetics of solubilization. In municipal sludge, heavy metal solubilization could be further complicated by the various metal complexes formed with the organic fraction, not to mention the possibility of microbial uptake. Furthermore, the chemical and biological characteristics of sludges which define heavy metal distributions will vary with sludge type causing potential differences in metal extractabilities upon acid treatment.

Although the acidification of municipal sludge has seldom been proposed for the full scale recovery of metals in the course of sludge treatment, extraction into 0.5 M acetic acid is a standard agronomic test applied to soils and other materials to determine the amount of heavy metal likely to be available for plant uptake (93). This kind of extraction was applied in the laboratory by Bloomfield and Pruden (93) to untreated municipal sludge that was incubated for three months in an anaerobic environment then aerated for three months. Anaerobic incubation of sludge resulted in decreased levels of acetic acid-extractable Cu, Zn, Cd, and Ni, whereas Pb extractability increased. Subsequent aerobic incubation, however, produced significant increases in extractability for all five metals.

A series of preliminary experiments were conducted at Cornell University to determine the potential of acid extraction to remove cadmium and other heavy metals from the following sludges: (1) primary and waste activated untreated sludge (21°C); (2) anaerobic digester effluent (35°C); and (3) autoheated aerobic thermophilic digester effluent (55°C). An aliquot of each sludge type was acidified with nitric acid to a pH of 2.0, continuously mixed and maintained at the temperature of the process from which the sample was taken. Samples from each acidified aliquot were periodically taken, centrifuged (18,000 x g), and filtered through glass fiber filters ($\leq$1.0 mm effective pore size), the filtrates were subsequently analyzed by atomic absorption spectroscopy for cadmium, chromium, copper, lead, nickel, and zinc. Heavy metal concentrations in these acidic filtrates were assumed to represent the solubilized metal fraction.

The results of the solubilization of cadmium from the raw, aerobic and anaerobic sludges are graphed in Figure 77. The most rapid release of cadmium



TOTAL Cd. CONTENT WET SLUDGE

| sludge | Cd. conc. mg/l |
|---|---|
| raw | 2.1 |
| aerobic | 4.4 |
| anaerobic | 4.7 |

Figure 77. Cadmium solubilization from raw, thermophilic aerobic, and mesophilic anaerobic sludges acidified to a pH of 2.0.

was observed with the acidification to a pH of 2.0 of sludge pretreated with an autoheated aerobic thermophilic digester. Virtually all of the cadmium was solubilized within six hours. The rates of cadmium released in the raw and anaerobically digested sludges, on the other hand, were far lower, with less than 20 percent of the cadmium in either sludge-type being solubilized in the same six-hour time span. More than 15 hours of acid treatment were required to dissolve even half of the cadmium in the raw and anaerobic sludges and quantitative solubilization with either sludge was never attained in the 24-hour test period. Rapid heavy metal solubilization was also observed with nickel and zinc upon acidification of the aerobic sludge. Copper, lead, and chromium were rendered soluble at lower rates. In all cases, the raw and anaerobic sludges exhibited a far lower rate of metals solubilization than did the thermophilic aerobically digested sludge as seen from comparing the percentages of metals dissolved in sludges after 12 hours of acidification, as shown in Table 28. Similar experimental results were obtained when hydrochloric acid was used in the metals stripping procedure in place of the nitric acid. These experimental results were repeated five times over a six-month operational period at the bench scale level (10).

TABLE 28. PERCENTAGE OF HEAVY METALS SOLUBILIZED IN RAW, MESOPHILIC ANAEROBIC, AND THERMOPHILIC AEROBIC SLUDGES ACIDIFIED WITH NITRIC ACID (pH 2.0) FOR 12 HOURS

| | Heavy Metals (percent solubilized) | | | | | |
|---|---|---|---|---|---|---|
| Sludge Type | Cd | Cr | Cu | Ni | Pb | Zn |
| Raw (21°C) | 51 | 32 | 0.2 | 72 | 37 | 78 |
| Anaerobic (35°C) | 39 | 6.6 | 0.1 | 81 | 47 | 76 |
| Aerobic (55°C) | 107* | 39 | 67 | 106* | 63 | 106* |

*Quantitative solubilization in 6 hours.

COMPUTER MODEL: RESULTS OF PREDICTION PROGRAM

Comprehensive thermochemical and kinetic equations describing the full scale autoheated aerobic digestion system under steady state operation were developed in an earlier section of this report. From this equation development came an equation for the prediction of reactor temperature as shown by Equation (38) below:

$$PT_R = \frac{HRT\ K_T\ BCOD_e\ R}{D_L\ C_{PL}} - [0.0407\ (T_R - T_A)^{1.2765}]\ HRT + T_O \qquad (38)$$

where $PT_R$ = predicted reactor temperature, °C

$HRT$ = hydraulic retention time, day

$K_T$ = reaction rate coefficient, $day^{-1}$

$BCOD_e$ = biodegradable COD in the effluent, gm/ℓ

$R$ = heat released during microbial respiration, Kcal/gm COD

$D_L$ = density of the liquid sludge, gm/ℓ

$C_{PL}$ = specific heat capacity of the liquid sludge, cal/gm-°C

$T_R$ = reactor temperature, °C

$T_A$ = ambient temperature, °C

$T_O$ = feed sludge, temperature, °C

The solution to the above equation for reactor temperature is a trial and error method, since both $PT_R$ and $T_R$ are unknowns. This kind of equation lends itself to a computer solution with an iterative step that eliminates the tedious problem of narrowing the difference between $P_{TR}$ and $T_R$ to a fractional number by manual calculation methods. The computer program also provided a wealth of other information useful in determining digester performance under given conditions. Variables in the computer program were $K_T$, R, $BCOD_O$, $T_A$, $T_O$, $C_{PL}$, and $D_L$. By using the thermochemical and kinetic equations, the computer model estimated the reactor temperature, BCOD loading rate, BCOD treatment efficiency, biological heat production, BCOD effluent, and the BCOD removal rate for hydraulic retention times of one to ten days. The program itself is given in Table 29, and an example of the printout information is given in Table 30. Note that for each printout table the variables were defined; therefore, a large number of data tables were generated since many values for each of the variables were investigated.

The computer model was one dimensional in that all variables were defined prior to the running of the program; however, the information that the program did produce was useful in illustrating the autoheating limitations (system specific, because of the empirical nature of the heat loss determinations) as well as providing a comparative index for actual system performance. The data in Figures 78 and 79 illustrate the importance of the biodegradable fraction of sewage sludge in achieving thermophilic temperatures. The results from the computer program show 15 gm/ℓ of biodegradable COD with a $K_T$ of 0.8 $day^{-1}$ to be the lower limit for autoheating with the present full scale digestion system (assuming $T_O = 15°C$, $T_A = 15°C$, $D_L \cong 1.0$ gm/ℓ, $C_{PL} \cong$ 1.0 cal/gm-°C). It is estimated that with an adequate heat exchanger system, heat could be recovered from the effluent and used to preheat the influent sludge, thereby enabling sludges with smaller biodegradable fractions and/or reaction rates to be autoheated to thermophilic temperatures.

The rate at which the biodegradable substrate (COD) is oxidized plays a major role in determining the reactor temperature. This concept is illustrated in Figure 80. The reaction rate coefficient, $K_T$, in turn, is a function of the substrate biodegradability as well as the dissolved oxygen levels, which makes it a controlling factor in substrate removal as illustrated by Figure 81. The average results from the steady state determinations are also shown in Figure 81. The relative positions of these data points are indicative of the differences in substrate composition and dissolved oxygen levels between the three systems.

A comparison of the results of steady state performance observed for the full scale system with the computer model is presented in Table 31. The results from the 3.20-day HRT compare favorably with the model results, as do the results from the 4.56-day HRT. The differences between the other HRTs and the

TABLE 29. COMPUTER PROGRAM FOR PREDICTION OF AUTOHEATED REACTOR TEMPERATURE

```
              REAL KT(15), LR, R(15), BCODO(18)
              READ, TA, TO, CPL, DL, II, JJ
              DO 100 K=1, II
              READ, KT(K), R(K)
      100     Continue
              DO 200 L=1, JJ
              READ, BCODO(L)
      200     Continue
              DO 10 M-1, JJ
              DO 10 J-1, II
              WRITE (6.20) KT(J), BCODO(M), R(J), TA, TO, CPL, DL
       20     FORMAT ('0', T5, 'KT=1', F5.3/T5, 'BCODO=', F6.3/T5, 'R=',
              F6.3/T5, 'TA=C', F6.3/T5, 'TO=', F6.3/T5, 'CPL=', F5.3/T5,
              'DL=', F5.3/T9, 'SRT', T19, C 'TR', 130, 'LR', T40, 'BCODE',
              T52, 'RR', T63, 'RE', T73, 'PTR', T84, 'BHP')
              HRT = 0.0
              DO 10 I-1, 10
              HRT = HRT + 1.
              BCODE = BCODO(M)/(1. + KT(J)*HRT)
              BHP = R(J)*KT(J)*BCODE
              RE = ((BCODO(M)-BCODE)/BCODO(M))*100
              LR = BCODO(M)/HRT
              RR = (BCODO(M)-BCODE)/HRT
              W = (HRT*KT(J)*R(J)*BCODE/(CPL*DL)+TO)
              A = 70.
              TR = 15.
              GO TO 11
       40     A = A/2.
              TR + TR-A
              GO TO 11
       41     A = A/2.
              TR = TR+A
       11     PTR = W-HRT*.0407*(TR-TA)**1.2765
              X = ABS(PTR-TR)
              IF (X-.3)15, 15, 12
       12     IF (PTR-TR)40, 15, 41
       15     WRITE (6.21)HRT, TR, LR, BCODE, RR, RE, PTR, BHP
       21     FORMAT ('0', 8F11.2)
       10     Continue
              Stop
              End
```

*WATFIV CARD ASSUMED

**JOB PAGES = 50

TABLE 30. COMPUTER PROGRAM PRINTOUT EXAMPLE, WHERE SYMBOLS ARE AS DEFINED EARLIER

| HRT | TR | LR | BCODE | RR | RE | PTR | BHP |
|---|---|---|---|---|---|---|---|
| 1.00 | 62.03 | 45.00 | 30.00 | 15.00 | 33.33 | 61.95 | 52.50 |
| 2.00 | 77.89 | 22.50 | 22.50 | 11.25 | 50.00 | 77.66 | 39.38 |
| 3.00 | 82.81 | 15.00 | 18.00 | 9.00 | 60.00 | 82.93 | 31.50 |
| 4.00 | 83.91 | 11.25 | 15.00 | 7.50 | 66.67 | 83.84 | 26.25 |
| 5.00 | 83.09 | 9.00 | 12.86 | 6.43 | 71.43 | 82.99 | 22.50 |
| 6.00 | 81.45 | 7.50 | 11.25 | 5.63 | 75.00 | 81.35 | 19.69 |
| 7.00 | 79.53 | 6.43 | 10.00 | 5.00 | 77.78 | 79.31 | 17.50 |
| 8.00 | 77.34 | 5.63 | 9.00 | 4.50 | 80.00 | 77.36 | 15.75 |
| 9.00 | 75.29 | 5.00 | 8.18 | 4.09 | 81.82 | 75.26 | 14.32 |
| 10.00 | 73.24 | 4.50 | 7.50 | 3.75 | 83.33 | 73.32 | 13.13 |

KT = 0.500
BCODO = 45.000
R = 3.500
TA = 15.000
TO = 15.000
CPL = 1.000
DL = 1.000

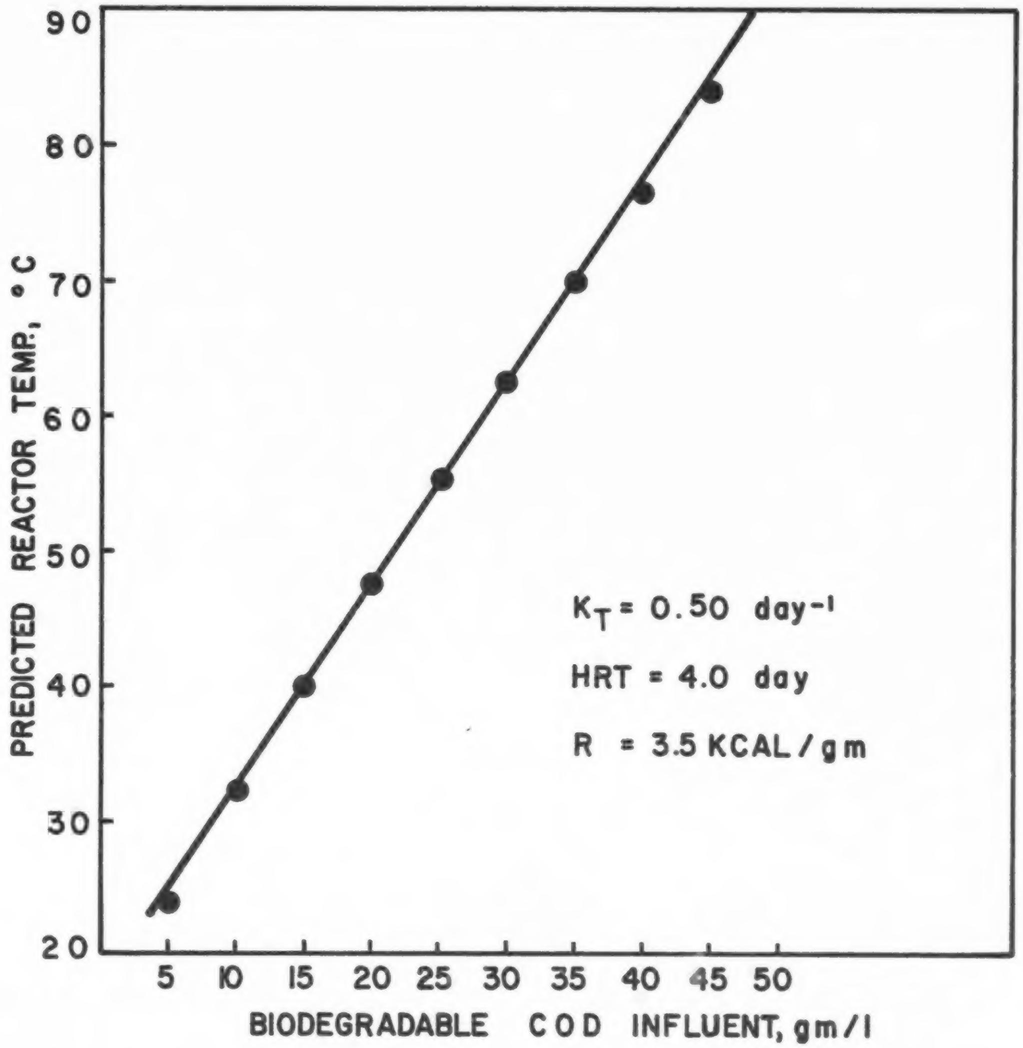


Figure 78. The influence of biodegradable substrate in the influent sludge on the predicted reactor temperature as given by the computer model.

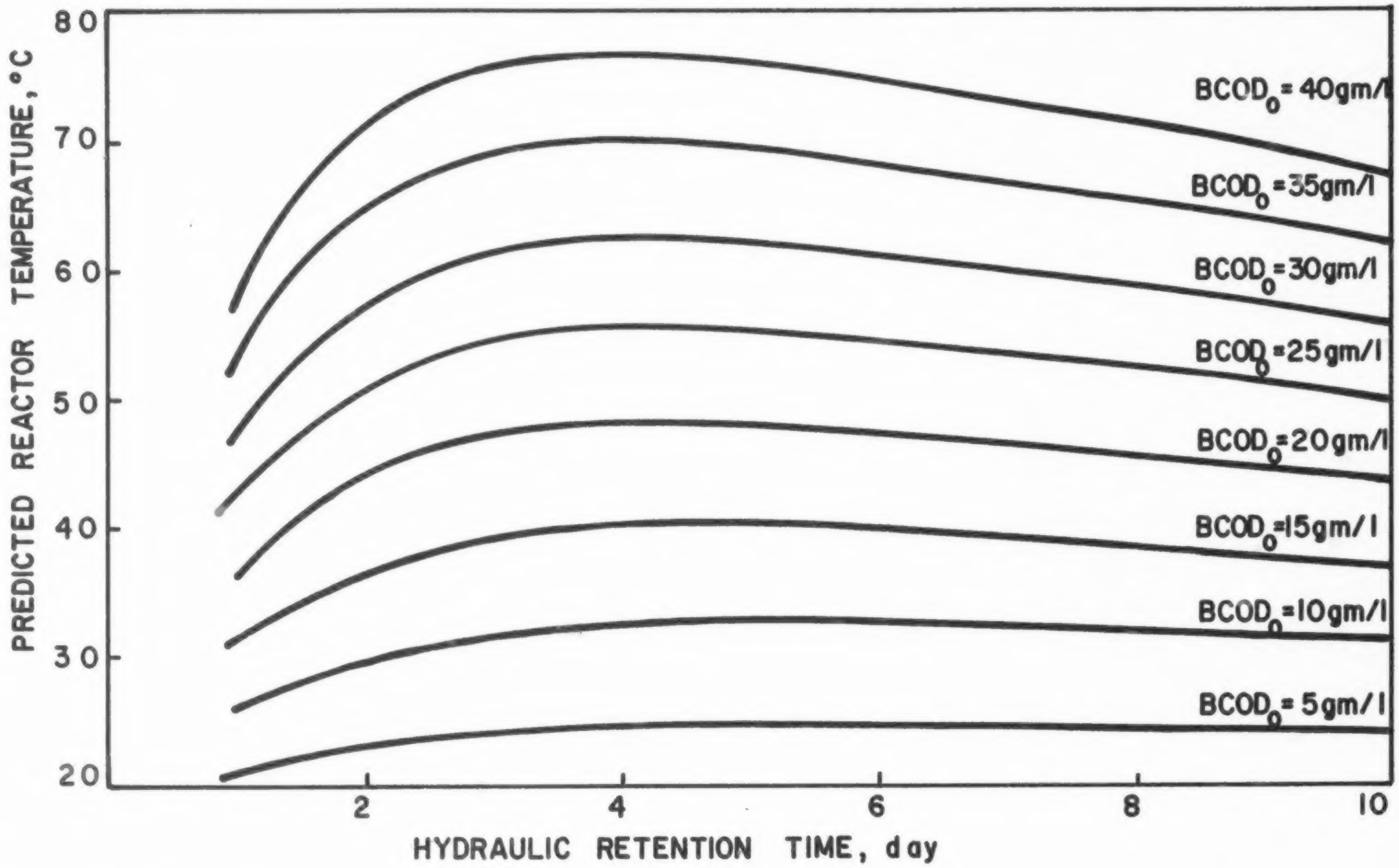


Figure 79. The influence of biodegradable organics (BCOD) on reactor temperature at various retention times (HRT) assuming $K_T$ = 0.4 $day^{-1}$ and R = 3.5 Kcal/gm with curves generated by computer model.

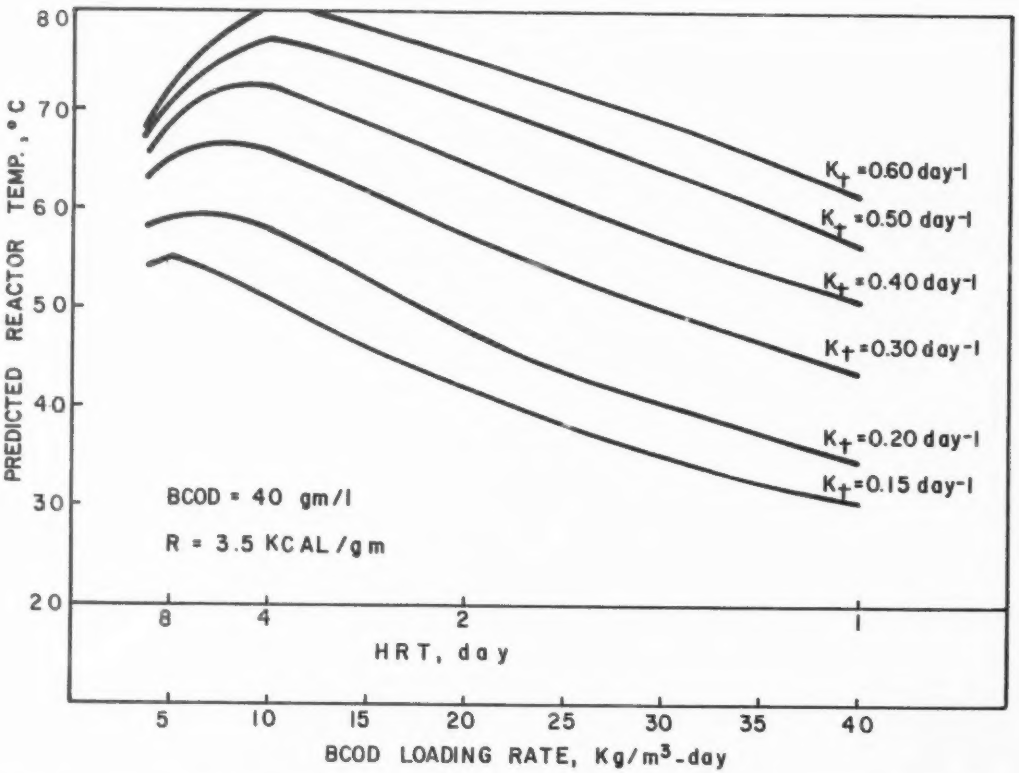


Figure 80. The reactor temperature shown as both a function of biodegradable loading rate and reaction rate as predicted by the computer model.

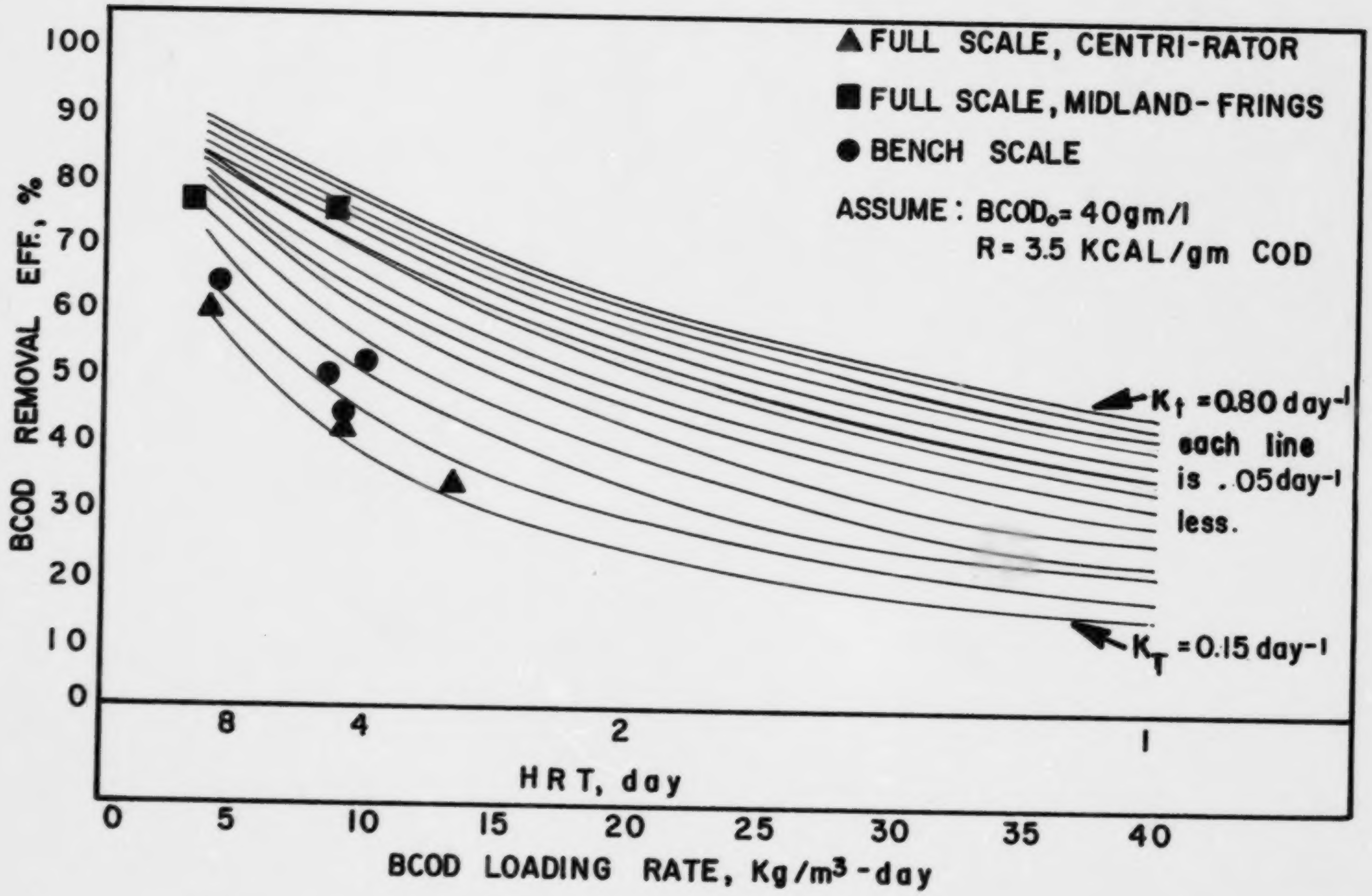


Figure 81. The relationship between treatment efficiency and loading rate for various values of $K_T$ as predicted by the computer model with all real steady state (average) values also shown.

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corresponding computer model results can be attributed to variations in substrate biodegradability, dissolved oxygen limitation, and inadequacies in the thermal loss equations (as discussed previously). This subject will be discussed in a subsequent section of the report.

TABLE 31. COMPARISON OF STEADY STATE PERFORMANCE WITH COMPLETE MODEL RESULTS

| | Full(C) | Full(M) | Full(C) | Full(C) | Full(M) |
|---|---|---|---|---|---|
| OBSERVED | | | | | |
| HRT (day) | 3.20 | 3.49 | 4.56 | 7.74 | 11.0 |
| Temp. Differential $T_R$-$T_o$ (°C)* | 38.2 | 41.8 | 41.7 | 25.1 | 50.4 |
| Temp. Differential $T_R$-$T_A$ (°C)* | 35.5 | 44.1 | 47.0 | 27.7 | 55.5 |
| $K_T$ ($day^{-1}$) | 0.16 | 0.81 | 0.15 | 0.20 | 0.30 |
| Biodegradable COD Removal Efficiency (%) | 34.1 | 74.0 | 41.1 | 60.7 | 76.8 |
| COMPUTER PREDICTION MODEL+ | | | | | |
| Approximate HRT (day) | 3.2 | 3.5 | 4.5 | 7.8 | 10 |
| Temp. Differential $T_R$-$T_o$ (°C)* | 37.6 | 54.0 | 37.4 | 34.5 | 42.9 |
| Temp. Differential $T_R$-$T_A$ (°C)* | 37.6 | 54.0 | 37.4 | 34.5 | 42.9 |
| $K_T$ ($day^{-1}$) | 0.15 | 0.80 | 0.15 | 0.20 | 0.30 |
| Biodegradable COD Removal Efficiency (%) | 32.3 | 73.4 | 40.2 | 59.6 | 75.0 |

*$T_R$ = reactor temp.; $T_o$ = feed sludge temp.; $T_A$ = ambient temp.

+Assumes R constant at 3.5 Kcal/gm COD. Note that results selected from computer tables that corresponded to actual results in terms of $BCOD_O$ and $K_T$.

Although the figures presented in this section indicated that reactor temperatures (as predicted by the computer model) could exceed 70°C, it is doubtful that the reactor temperature could be maintained above 70°C even with heat exchangers for recovering heat from the effluent. It is estimated that temperature and/or oxygen availability would be limiting factors in the growth of the thermophilic microorganisms, thereby limiting system performance in terms of reactor temperature and substrate removal.

## OPERATIONAL EXPERIENCES AND PROBLEM IDENTIFICATION

One of the primary objectives of this study was the documentation of the practical operation and maintenance requirements of a full scale thermophilic

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aerobic digestion process. Surprisingly, few operational problems were experienced with the process itself. Although the weather conditions and the size of the experimental equipment created demanding situations and problems of their own, e.g., freezing of exposed valves, no significant problems arose in maintaining the reactor temperature or other components associated with the digestion process. The following section details the general operational and maintenance requirements, and the few problems encountered during the eighteen months of full scale thermophilic aerobic digester operation. A summary of the major problems experienced is presented in Table 32.

TABLE 32. PROBLEMS EXPERIENCED WITH FULL SCALE SYSTEM AND ASSOCIATED DATES

| Date(s) | Problem |
|---|---|
| 6/01/77 - 6/17/77 | -Centrifugal pump, Crisafulli model CPO2, was found to be underdesigned with a 1.5 kW (single phase, 230 volt) electric motor and was subsequently replaced with a larger motor (3.7 kW, 3 phase, 230 volt). |
| 7/01/77 - 7/21/77 | -Foaming problems; new foam cutter blade installed. |
| 12/15/77 - 2/03/78 | -Freezing problems. |
| 2/16/78 | -Rags plugging feed pump; bearings on pump lubricated. |
| 3/25/78 | -Rags plugging feed pump; new bearings installed and impeller trued. |
| 6/01/78 | -Rags plugging feed pump; new bearings and lower drive shaft installed. |
| 6/13/78 | -Rags plugging Midland-Frings aerator. |
| 11/01/78 | -Grit accumulation in reactor due to malfunction of existing grit chambers. |

## General Operation and Maintenance Requirements

The full scale system required relatively little manpower input on a daily basis. Approximately one to two man-hours per day were required for operation of the system on a continuous flow basis at an intermediate loading rate. The following list of tasks composed the routine (daily) work on the system:

(1) Monitoring of the feed tank depth for determination of hydraulic flow rates and determination of the need for addition of feed sludge to the feed tank.

(2) Filling of the feed tank every few days as determined above in (1).

(3) Monitoring of airflow, amperage, and power, on the aerator for determination of aerator performance.

(4) Monitoring of effluent tank depth as a check on hydraulic flow rates.

(5) Emptying of the effluent tank at the end of a hydraulic retention period.

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EPA-287 (Cin.)
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TYPING GUIDE SHEET

(6) Measurement of dissolved oxygen levels in the reactor.

Of course, more time was needed on sampling dates which occurred at the end of a test hydraulic retention period. It was estimated that daily man-time could be reduced if the aerobic digester was fed thickened sludge in a manner similar to conventional anaerobic digesters, i.e., feed supplied on a timed basis by metered piston pumps connected directly to the thickener(s) cone outlet. This procedure would have eliminated the need for a feed tank, but would not have been sufficient for the study requirements since measurements to construct mass balance were necessary.

It is pertinent to note here that under actual continuous operation of the autoheated thermophilic aerobic digestion process (real world application), the daily operational requirements could be integrated into the daily routine of the treatment plant operators. The system is simple to understand and operate, unlike anaerobic digesters which require an understanding of a complex gas collection and reactor heating systems.

## Weather Associated Problems

The full scale system used in this study was, of course, an experimental system; therefore, all of the equipment was placed with the idea that the entire system would be temporary, i.e., it would be operated for two years only. This meant that all of the equipment was located above ground. Due to the severity of the winter weather in the northeastern United States, this produced freezing problems for exposed valves and pipes. A delay in second year funding prevented the complete winterization (see a previous section) before some freezing problems occurred.

The delay in funding, however, was useful in that it provided an insight into the magnitude of weather-associated problems. The cold weather problems could be avoided by locating the system in-ground with all pipes and valves either below frostline or in a housing structure for ease of access. A subsurface location or a completely enclosed superstructure would be a necessity for a system designed for operation at a sewage treatment plant located in a temperate climate.

## Equipment Problems

The autoheated aerobic digestion system used the following five large pieces of motorized equipment:

(1) centrifugal, vertical drive pump (Crisafulli model CP02)
(2) mechanical foam cutter (DeLaval)
(3) centri-rator aerator (DeLaval)
(4) Midland-Frings aerator (LFE Corp.)
(5) centrifugal solids pump (Hydromatic)

Each of the above pieces of equipment experienced at least some minor mechanical problems. These problems, their solutions, and the implications of these problems as concerns the operation of the system are presented here.

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Centrifugal Feed Pump--
The majority of the problems experienced for the entire system were associated with the feed pump. This piece of equipment was responsible for most of the down time for the system. Due to the harshness of the environment, i.e., thickened primary and secondary sludge that was abrasive and corrosive, and the fact that the pump drive shaft was not enclosed (see Figure 82), the life of the four main bearings was limited to approximately four to five months of continuous operation. The thrust bearing (see Figure 82) was especially susceptible to failure. Its failure was due in part to the abrasive nature of the sludge, but the prime reason for failure was the plugging of the impeller housing with trash which forced the bearings out of the race, thereby leading to pump failure. All of the shaft bearings were replaced three times and during each replacement procedure the impeller was cleaned and trued. Approximately sixteen man-hours were required for this task.

On two separate occasions, the Binghamton-Johnson City STP experienced a large influx of rag material which caused malfunction of the plant comminutors. The rags that escaped the comminutors caused plugging problems with the feed pump. On these occasions, the pump was removed from the feed tank, cleaned (rags removed from the shaft and impeller housing), and returned to the feed tank. Approximately four man-hours were required for this procedure.

The above described problems could be eliminated with the use of a timed and metered positive displacement (piston) pump such as those traditionally used for pumping sludge from thickener cones to anaerobic digesters.

Mechanical Foam Cutter--
On only one occasion did the mechanical foam cutter fail to contain the foam produced in the reactor. It was thought the excessive amount of foam was either induced microbially from shifting microbial populations during a change in reactor loading conditions, or chemically by a large influx of surfactants to the treatment plant. A larger blade was installed on the foam cutter shaft which was able to control foaming during the rest of this study. (Note that the new foam cutter blade was installed during the first two months of operation and performed adequately thereafter.)

Centri-rator Aerator--
The rag problem discussed earlier had little effect on the centri-rator. However, the abrasive environment did have an attritional effect on the impeller (see Figure 83). It was estimated the impeller would need replacement every two to three years. For proper operation, the shaft bushings (see Figure 84) on the centri-rator were adjusted approximately every five months. This adjustment also acted to extend motor and motor bearing life. Note that a diagram of the centri-rator is given in Figure 84.

The float system, upon which the centri-rator was suspended, experienced a buildup of sludge on the styrofoam floats. This sludge layer caused a change in the buoyancy of the floats, thereby lowering the impeller immersion depth. Although the consequence(s) of this buildup was not studied, it was estimated that the sludge accumulation should be removed every few months. The float system, i.e., the styrofoam blocks, deteriorated rapidly (see Figure 85). It was estimated that new styrofoam blocks would be needed every twelve months.

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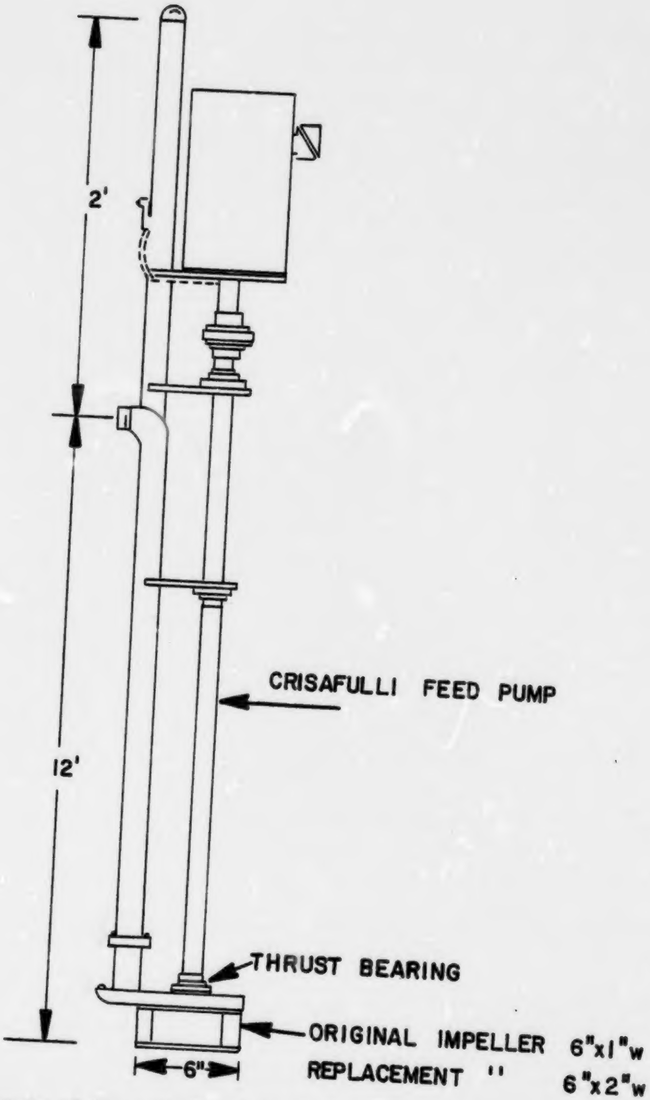
2'
12'
CRISAFULLI FEED PUMP
THRUST BEARING
ORIGINAL IMPELLER 6"x1"w
REPLACEMENT " 6"x2"w
6"

Figure 82. Diagram of the centrifugal feed pump illustrating the various components.

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Figure 83: Photograph of centri-rator impeller.

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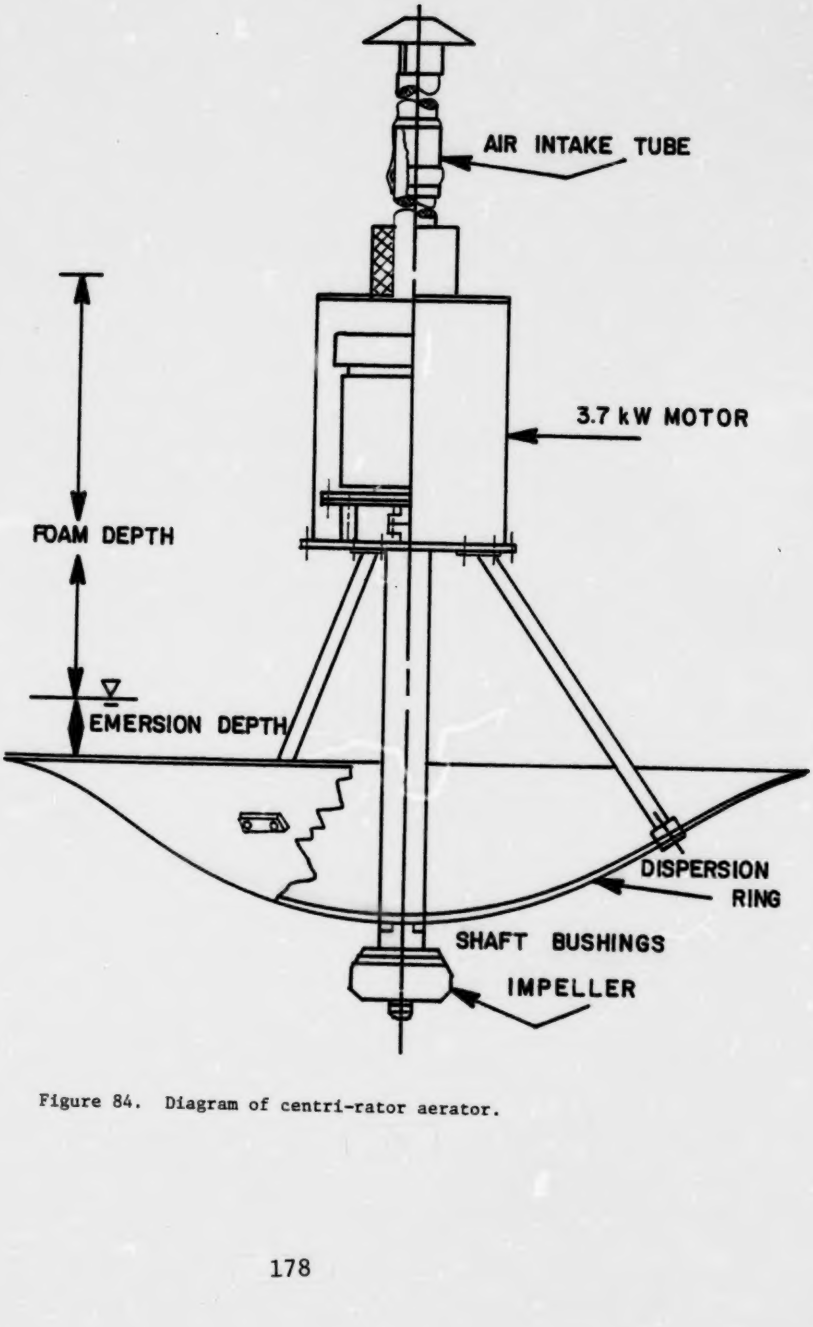
AIR INTAKE TUBE
3.7 kW MOTOR
FOAM DEPTH
EMERSION DEPTH
DISPERSION RING
SHAFT BUSHINGS
IMPELLER

Figure 84. Diagram of centri-rator aerator.

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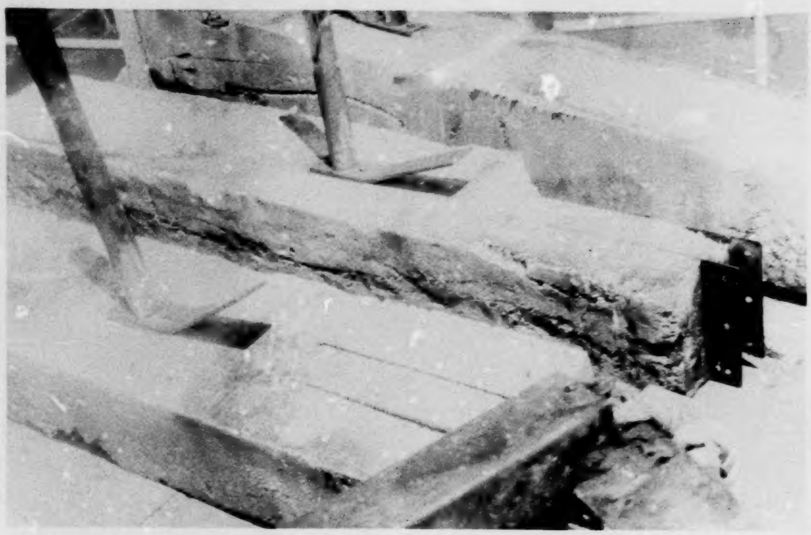

Figure 85: Centri-rator styrofoam float system illustrating deterioration of the styrofoam blocks.

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EPA-287 (Cin.)
(4-76)

TYPING GUIDE SHEET

The deterioration of the styrofoam appeared to be a function of the high temperature and chemical nature of the sludge.

Midland-Frings Aerator--
The problems associated with the Midland-Frings were primarily due to the vast amount of rag influxes mentioned earlier. The stator plates that encased the impeller enabled rag debris to become entrapped in the sludge intake area as illustrated by the photograph in Figure 86. This rag accumulation inhibited the rotation of the impeller, thereby overloading the aerator motor which inturn was deactivated by thermal-sensitive (overload) breaker switches. This action protected the motor from damage; however, this action also interrupted aeration of the reactor contents. On two separate occasions, steady state operation was completely interrupted. The aerator was then removed and the stator plates separated for cleaning of the impeller. Approximately four man-hours were required for this cleaning procedure and the subsequent reinstallation of the unit.

On one occasion, the performance of the Midland-Frings unit was hampered by an unusually large amount of grit, i.e., inorganic solids, in the feed sludge. The inorganic solids, due to their high specific gravity, formed a thick sediment layer in the reactor, thereby limiting the mixing ability of the aerator. Malfunction of the grit chambers was responsible for the large amount of grit in the sludge. The reactor contents were diluted with water to suspend the accumulated grit and allow the grit to flow to the effluent tank by way of the gravity overflow when the tank volume was exceeded by the addition of the water. Once the grit layer was removed, continuous operation was initiated.

## Odor Production

On several occasions, during humid summer weather, complaints arose concerning the odor apparently eminating from the stack gases of the thermophilic reactor. The odor of the stack gas was similar to a properly operating activated sludge system, in that an actinomycetes' odor was pervasive; however, the gas also smelled of hot ammonia. This was to be expected, because ammonia was volatilized during aerobic thermophilic treatment as shown by the steady state operational data. Some people found this odor extremely objectionable, while some people hardly noticed an odor. The problem with ammonia odors was noted in a study on thermophilic aerobic digestion of swine manure in Germany (64).

It was estimated that the ammonia could be removed by scrubbing the gas with an acidic solution of water. This would control odors originating from the hot ammonia gas, with low level technology and cost.

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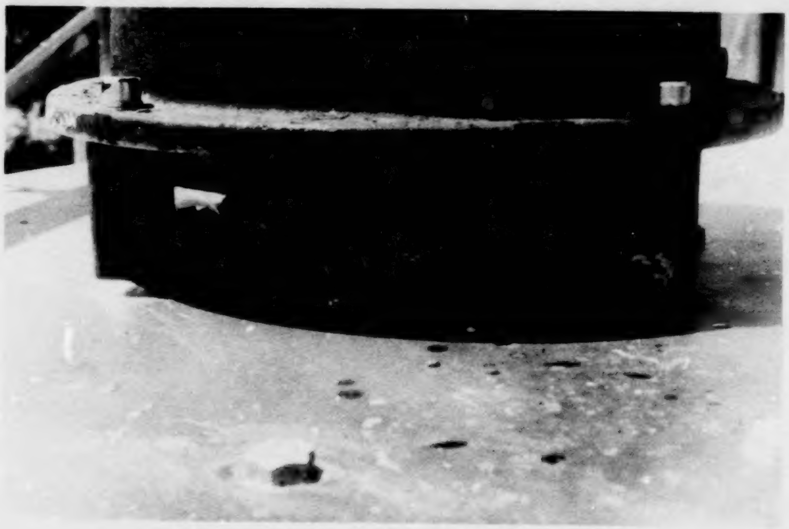

Figure 86: Midland-Frings aerator with accumulated rag debris in sludge intake area.

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## SECTION 8

## DISCUSSION OF EXPERIMENTAL RESULTS

The results of this two-year study indicate that autoheating during aerobic digestion can be controlled to achieve effective solids treatment using air aeration. Sludge processing rates exceeding that produced by a population of 10,000 people were tested for long-term periods under real world conditions. All objectives of this study were achieved, and the successful results of this first large scale test indicate that the technology is a cost effective alternative to conventional sludge treatment.

### PRACTICAL OPERATIONAL PROBLEMS

Practical problems that were considered to be potential problem areas included start up problems in terms of time, availability of well balanced microbial population capable of operation at high temperatures, and the impact of high temperatures on the equipment. Three years of operation of similar equipment in an earlier study with animal wastes indicated that the practical operational problems were minimal (6,7). This was also true of this study with sewage sludge.

Start up and autoheating of liquid sludge to temperatures between 45°C and 66°C occurred immediately and with no special precautions. This occurred even under adverse winter conditions when feed sludge temperatures approached freezing and the air temperature was far below 0°C. This indicated that the thermophilic microorganisms capable of growth at high temperatures were always present in the feed sludge. Reasons for the ubiquitous nature of this group of organisms are not known. However, rapid start up and acclimation of this process does not appear to be a negative consideration.

Nearly all of the operational problems were associated with the temporary nature of a test facility designed for measurement of mass balances. For example, exposed pipes and valves for sampling, feed, and wasting froze during the first cold season operation. Precautions to prevent this in an actual full scale facility would have been taken.

It should be noted that the long term testing was accomplished with the actual sludge generated by a facility subjected to considerable variations in loading and flow rates (see Section 5). An example of an extreme problem that would occur in real world operation was the clogging of the Midland-Frings unit with large rags. This occurred during a period when a major spring sewer cleaning operation was on-going. The dose of rags that was introduced into the facility took a number of components of the 69,000 $m^3$/day plant out of order, such as the comminutors. When the rags reached the sludge treatment

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facilities, they were highly concentrated. Nearly 1.5 feet of rags accumulated in the bottom of the reactor and the intake to the aerator became clogged. This overloaded the electrical circuits, and it turned itself off by activating the breaker before permanent damage occurred. The unit was placed back into operation within 48 hours after the unit was taken from the tank, cleaned, and replaced.

## AUTOHEATING AND AERATION REQUIREMENTS

As noted in the Theoretical Section earlier, studies have emphasized the need for an effective aeration system to minimize the amount of energy lost through evaporation in order to achieve autoheating with the small amount of energy available in liquid sewage sludge. These studies have often concluded that pure oxygen would be one alternative that would minimize energy losses and achieve autothermal conditions. This study showed that the simple self-aspirating aerators achieved exceptionally high oxygen transfer under adverse conditions, and this resulted in a decrease of energy lost as evaporation so that autoheating was successful. These aerators require no compressors or blowers and are operationally as simple as most floating aerators. Thus, they offer an alternative method of conserving the energy in the aeration process.

Comprehensive efforts to characterize these aerators in tap water, and to relate these operational characteristics to corresponding conditions in the waste, were attempted using conventional methods. Perhaps one of the important observations of this study was the fact that none of the conventional methods of characterization of aeration were applicable. High solids concentrations and temperatures enhanced oxygen transfer rather than decreasing it as would be expected. In fact, based on earlier studies, i.e., the results of the tap water tests showing average transfer efficiencies of 2 percent at 60°C, it would have been concluded that the two aerators tested could not have achieved the oxygen transfer efficiencies required to minimize evaporation losses to enable energy conservation during aeration.

Direct observations clearly indicated why the aerators achieved high transfer efficiencies in sludge. The resulting bubbles entrained in the slurry were extremely small, with the majority in the 100 to 150 micron diameter size range, thus causing a very large surface area through which oxygen transfer could occur. Because of the small size, it is likely that the average bubble residence time was long because of its tendency to move with the direction of the liquid. It can only be speculated that the effect of solids and dissolved salts in the sludge caused a significant reduction in bubble size over that which was formed in tap water at the same temperature. Additional studies on this topic might lead to significant increases in the understanding of the theory of aeration. Studies in this area are continuing at Cornell University.

The autoheated characteristics of this system depend upon a number of parameters, one of which is a well insulated reactor. During aeration, energy losses from the reactor (when operated at an air-water temperature differential

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of about 40°C) were 4.8°C per day, or about 4.8 Kcal per liter per day, resulting in a total loss of energy of 136,320 Kcal per day. This equals an average loss per unit surface area of 1,726 Kcal per $m^2$ per day (includes heat losses through aeration). The cooling energy losses can be compared to the biological energy generation by reviewing the loading rate relationships to removal efficiencies (Figures 52,59,60). Typical organic removal rates of 4 to 6 kg COD per $m^3$ per day resulted in biological energy generation rates of 14 to 21 Kcal per liter per day (see Figure 61), or a total reactor generation rate of 497,000 Kcal per day. This has a fuel replacement annual value of about $2,700 for the small reactor. Thus, the heat losses through the walls of the reactor and through aeration accounted for about 27 percent of the total energy generated under these conditions. It should be emphasized that the heat losses in tap water were only estimates of comparable losses from aerating slurries. Because of the heat generation characteristics of the slurry, energy losses from the slurry could not be measured directly. It is likely that the thick foam layer and the bubble size caused significant differences in heat loss over that measured with tap water. In all cases, the tap water losses were assumed to be equal to the slurry losses.

The other major energy requirement that must be satisfied is the energy to raise the sludge temperature from the ambient level. If the influent slurry temperature is 15°C, the energy required to raise the temperature to 50°C would be 35 Kcal per liter. Thus, if the net energy available for heating were 21 Kcal per liter of reactor per day, and 27 percent were lost, it would take an average residence time of 2.3 days to maintain the reactor at 50°C. This activity level would correlate to a total volatile solids destruction of 10 gm TVS/ℓ. In other words, if the quantity of biodegradable organic matter were available, the reactor would achieve autoheating to 50°C with the air-aerators tested in this study.

A computer program and the wide range of variables tested in this study indicate the conditions that will enable autoheating to occur. Of course, the data developed here are system specific and should be used only as a guide for other designs. Larger reactors, for example, may result in lower energy losses, thus enabling higher operating temperatures to be achieved. Because of the large number of variables involved, it is difficult to predict the operational conditions without using a computer analysis with a significant number of assumptions for input into a model. However, as can be seen from the previously discussed simplistic example of an input sludge with a solids content of 3.8 percent dry matter (assuming that 40 percent of the TVS was oxidized in 2.3 days, and that the influent sludge had a TVS content of 65 percent of the TS), it is likely that sludge with a total solids content as low as about 2.5 percent would autoheat to thermophilic temperatures. In cases where a sludge cannot be thickened to this concentration, the use of a heat exchanger on the effluent would enable high slurry temperatures to result. Thus, with this addition to a reactor, it can be concluded that nearly all sewage sludges can be treated with this system at thermophilic temperatures without the addition of external energy sources, providing that the sludge has at least a 40 percent TVS biodegradability.

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## AERATOR PERFORMANCE

The two aerators tested had similar oxygen transfer efficiencies with maximum measured values in excess of 20 percent. Since both aerators had a fixed aeration capacity, an increase in loading rate would eventually add an oxygen demand that would exceed the transfer capability of the aerator(s). Note that the two units are not directly comparable, since the Midland-Frings unit had a 5.22 kW (7 hp) rating and the centri-rator a 3.73 kW (5 hp) rating. Visual comparisons indicated that the bottom mounted Midland-Frings unit had a great deal more mixing capability than the liquid surface mounted centri-rator. Each unit delivered between 24 and 40 $\ell$/sec of air to the unit or about 97 $m^3$ per $m^3$ reactor per day (97 $ft^3/ft^3$ reactor-day). This is slightly higher than the aerobic digester design requirement of about 90 $m^3/m^3$ reactor-day for combined primary and waste activated sludge.

Increased organic loading rates increased the oxygen transfer efficiency. A doubling in organic loading caused the observed oxygen transfer efficiency to increase from 14 to over 18 percent for the centri-rator, while the Midland-Frings unit increased from 8 to 22 percent, the highest value observed in this study. Although the influence of increasing temperature is not known, it is likely that the lowered dissolved oxygen content that occurred with increased loading rates (see Figures 64 and 65) was largely responsible for this increased efficiency.

The oxygen transfer data indicate that the power input should be closely matched with the oxygen demand so that dissolved oxygen levels are maintained at relatively low levels. This would result in maximum autoheated temperatures and minimum power requirements. Average reactor dissolved oxygen values just above zero were adequate to prevent odor formation. The best practical arrangements for maximum removal rate and minimum aeration requirements appear to include use of the Midland-Frings unit at an organic loading rate of around 15 kg of biodegradable COD per $m^3$ per day, with an aerator power input of 0.18 $kW/m^3$ of reactor.

## EFFECT OF TEMPERATURE ON REACTOR KINETICS

The average full scale reactor temperature varied from 47°C to 62°C (see Figure 50) depending upon the air and feed sludge temperatures, and the feed rate. The most extreme example of the system's capability for autoheating occurred late in the study when the reactor temperature reached 66°C, with the air temperature at 0°C, and the feed sludge at 12°C. Note that whenever operating conditions are compared in this study, the temperatures of operation should be included in the evaluation. In most areas, an attempt was made to clarify the effect of this parameter. Another complicating variable was the low operating dissolved oxygen (D.O.) levels that occurred at the high organic loading. In most cases, the D.O. levels in one-quarter to one-half the reactor volume varied from 1 mg/$\ell$ up to 6 mg/$\ell$, except for the highest loading rate (3.2 day HRT) which had a D.O. concentration less than 0.1 mg/$\ell$. Before the degree of oxygen limitation could be determined, the turnover rate of the reactor contents would have to be known. No information was obtained

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TYPING GUIDE SHEET

on this topic, except to note that the Midland-Frings was more efficient for mixing.

The influence of the loading rate on the resulting reactor temperature is not as significant as might be expected (see Figures 57 and 58). At low loading conditions, the maximum temperature may be limited by the amount of biodegradable organics, whereas at high loading rates, the quantity of energy required to heat the large through-put volume limits the reactor temperature. Typical temperature differentials between the reactor and the feed slurry were 30°C to 50°C. Peak temperature differentials occurred at loading rates around 9 kg/m$^3$-day for the centri-rator.

A combination of autoheated aerobic digestion followed by anaerobic digestion has been suggested as a system which could eliminate some of the operational problems associated with anaerobic digestion. In order to operate at 50°C, this would require a minimum autoheated temperature differential of 35°C. Such a differential would be created by oxidation of a quantity of solids of about 13 to 15 gm/ℓ of COD or about 10 gm/ℓ TVS. In sludge concentrations varying from 2 to 6 percent total solids, this represents between 20 and 50 percent TVS destruction efficiency. Thus, except with concentrated sludges, such a system would consume a large quantity of organics prior to anaerobic digestion. Applications of the anaerobic digestion process at this point would introduce a highly complex process for relatively few apparent advantages.

The influence of increasing reactor temperature on the removal rate of the biodegradable organics is shown in Figures 54 to 56. Increasing the reactor temperature from 45°C to 62°C appeared to more than triple the value of $K_T$, since it ranged from an average of about 0.18 day$^{-1}$ at 47°C to nearly 0.8 day$^{-1}$ at 62°C. This temperature effect is complicated, however, since the composition and biodegradability of the feed sludge varied over 100 percent during the study as was shown in the batch biodegradability tests (Table 15). The influence of plant operation on the variability of sludge composition was discussed earlier. However, it appeared that the doubling of the removal rate between 46°C and 50°C from 0.18 day$^{-1}$ to about 0.36 day$^{-1}$ was consistent for much of the data focused here. This effect of temperature on the removal rate coefficient agrees with others who have observed this effect during sewage sludge stabilization (14,21).

## Microbial Efficiency of Substrate Utilization

The quantity of heat lost to the environment during substrate oxidation was discussed in the Theoretical Section. The average value of heat released during oxidation is approximately constant and should be approximately 3.5 Kcal/gm COD oxidized. Careful accounting of energy flows enables one to determine its estimated value (R). Wright (19) in small scale experiments showed that R varied from a low of 3.01 Kcal/gm COD removed to a high of 4.36 Kcal/gm of COD removed, with the value increasing gradually in direct proportion to increasing retention time. It could be hypothesized that the exact value for R depends upon microbial cell yields per gram of substrate utilized, and that at short HRT values, a higher fraction of substrate is converted into new cells more efficiently and at longer HRT values the net

EPA-287 (Cin.)
(4-76)

cell synthesis is low per unit of substrate oxidized, thus resulting in a higher value of R. This hypothesis is supported by a review of energetics as proposed by McCarty (71). This pattern could not be shown in this study as shown in Figure 62 probably because of the larger experimental error involved in obtaining data to estimate R from the full scale operation.

## FATE OF NITROGEN DURING HIGH TEMPERATURE TREATMENT

Considerable energy savings might be incurred when thermophilic aerobic digestion is used over digestion (aerobic) at lower temperatures, since the nitrification reaction is inhibited altogether. However, high temperatures and pH values favor the stripping of ammonia gas from solution. Here again, the low gas volume used by the self-aspirating aerators was a benefit since only minor quantities of nitrogen were lost in the gas. In the small batch experiments, with HRTs ≥25 days, nearly all the nitrogen was lost from the system. Maximum nitrogen losses from the full scale and bench scale units, as noted in Table 16, ranged from 8 percent to 28 percent. The TKN was apparently air stripped as ammonia (after the organic nitrogen was degraded) from the system.

## DEWATERABILITY EFFECTS OF HIGH TEMPERATURE TREATMENT

Previous studies have indicated increased ease of dewaterability. In the long term (30 days or longer) batch laboratory experiments, the solids eventually settled rapidly to about 25 percent of their original volume and dewatered easily enough for possible application to sand beds. However, the effluent from the full scale system had a sludge with dewatering characteristics considerably worse than those of the raw sludge.

The decreased dewaterability observed with these sludges is not entirely unexpected (94). All high rate biological reactors, especially high temperature units, tend to develop a bacterial mass that is rapidly growing and ideal for creating difficulties in dewatering of the organic solids. This may be an important consideration in the ultimate process design, if dewatering the sludge is a facility requirement.

## PATHOGEN SURVIVAL

The fate of three groups of pathogenic organisms in the thermophilic aerobic digestion process and in the mesophilic anaerobic digestion process (both were full scale systems) was studied utilizing only the background levels of these organisms in the raw sludge. This kind of study (full scale with no organism addition) has been reported to produce data that are difficult to evaluate in comparison to small scale tightly controlled seeded reactors in terms of sampling and condition controls (52). However, it was felt that consistent data, with regard to the number of organisms recovered from each of the sludges, were indicative of reliable sampling and analysis procedures. It was realized that fluctuations in the background levels of pathogens could have been due to nonrepresentative samples or inadequacies in screening high solids sludge samples for pathogens.

### Bacteria

The performance of the autoheated aerobic thermophilic digester in the inactivation of pathogenic bacteria, viruses, and parasites was exemplary of the effect of a high temperature treatment process on pathogens. On the other hand, the conventional mesophilic anaerobic digester performance, with regard to pathogen inactivation, was typical of performance data reported in the literature.

The reduction of *Salmonella* sp. from feed concentrations as high as 2,200 organisms per 100 ml wet sludge to below the limits of detection in all of the aerobic thermophilic samples was a function of reactor temperature as well as the chemical nature of the sludge environment. In all but one sample, the reactor temperatures were ≥45°C. This temperature, combined with an exposure time of approximately 24 hours, was shown to effectively reduce *Salmonella* sp. to below detectable limits (60). The reactor temperature for the other sample date (June 26, 1978) was 40°C. Strauch (68) showed that an exposure time of 24 hours at a temperature of 40°C and pH of 7.8 was sufficient to reduce *Salmonella dublin* below detectable limits in aerated liquid swine manure. Strauch (68) also reported that inactivation of *S. enteritidis* in swine manure required 40 hours of exposure time at 42°C and pH 8.0. Therefore, it can be concluded that inactivation of *Salmonella* sp. in aerating sewage sludge requires temperatures above 40°C with the pH of the slurry and the species of *Salmonella* playing important roles.

Reductions of *Salmonella* sp. in the anaerobic digesters operated at 35°C were comparable to those reported in the literature (53), ranging from 44 to 99 percent. The inability of the anaerobic digesters to completely inactivate *Salmonella* sp. was a function of the low operating temperature of 35°C. Competitive inhibition of the *Salmonella* sp. by saprophytic bacteria probably played a role in keeping the number of *Salmonella* low.

The fates of the other indicator organisms, *Pseudomonas aeruginosa*, fecal coliforms, and fecal streptococci followed the same trends as *Salmonella* sp. in both the anaerobic and aerobic digesters. The larger percentage of organism reduction in the thermophilic aerobic digester was due to the higher temperatures of operation.

The Cornell University Veterinary Diagnostic Laboratory performed analyses for *Enterobacteriaceae* on three separate occasions with no significant findings. The inability of this laboratory to isolate *Salmonella* sp. was probably due to inexperience in screening high solids, i.e., sludge samples, whereas the U.S. EPA laboratory utilized techniques developed by that group for isolating and enumerating *Salmonella* sp. from various kinds of sludges.

### Viruses

Virus inactivation (Δ total plaque forming units) by the thermophilic aerobic digester was complete on all but one occasion. On that sample date, August 14, 1978, the digester was marginally operational. The complete inactivation, i.e., below limits of detection, of viruses at temperatures ≥40°C and pH of 7.0 was expected from the literature (52,57,58,61,62,63). This synergistic

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action of temperature and pH also worked effectively in reducing viruses in the anaerobic digester. Although viruses were not reduced below detectable limits, they were significantly reduced from the numbers in the raw sludge with destruction efficiencies as high as 98.6 percent. The loss of virus infectivity in both the anaerobic and aerobic digesters, as found by previous studies, was probably due to the cleavage of the viral protein coat followed by activation of the viral RNA (58).

## Parasites

The results of the parasite analyses were difficult to interpret, due to the low numbers of parasites recovered in all of the samples, especially those samples analyzed by the University of Cincinnati for *Ascaris* sp. The low number of parasites found in the raw sludge was a function of the climate and population of Binghamton, New York. Previous studies have shown high numbers of various parasites in raw sludge, e.g., ≥50,000 *Ascaris* eggs per kilogram dry sludge (87), with organism concentration dependent upon climate as well as on the socioeconomic structure of the community (65,87). When the data from Tulane University's parasite analyses were presented as total viable and non-viable parasitic ova, the aerobic thermophilic digester was shown to be superior to the anaerobic mesophilic digester in the reduction of viable parasitic ova (Table 26). The high reactor temperatures in the aerobic digester were responsible for the observed decrease in viable ova over the anaerobic unit. On two sample dates, no viable ova could be isolated from the aerobic digester sample, whereas nine non-viable ova were found on each of the two dates. Since a small number of viable ova were isolated from the aerobic digester on five other sample dates, it must be concluded from the (limited) available data that the number of parasitic ova are greatly reduced by the aerobic thermophilic digestion process. Complete inactivation of parasitic ova and cysts would require temperatures of ≥60°C with corresponding exposure times of greater than or equal to one hour. This could be accomplished in a multi-stage digestion system in conjunction with heat exchangers for boosting the reactor(s) temperature to around 70°C.

With regard to the various intestinal parasites isolated by the School of Tropical Medicine, Tulane University, *Toxocara* sp. were isolated with the greatest frequency. On most occasions, the aerobic digester inactivated these organisms, with the anaerobic digester showing a much lower level of inactivation. *Toxocara* sp. are known as the cat and dog ascarids. Their presence is indicative of the presence of feline and canine feces in the sludge. Human infestation by these organisms is possible, causing a disease not dissimilar to ordinary ascariasis. *Trichuris* sp. (*T. trichiura* and *T. vulpis*) were isolated with the next greatest frequency. On all but one occasion, these organisms were completely inactivated by the aerobic thermophilic digester, whereas the anaerobic unit exhibited very poor inactivation performance. *Trichuris* sp. are parasitic to man and other vertebrates, and are known as whipworms. The organism isolated with the third most frequency was *Ascaris* sp. These are roundworms of great economic importance, since they may infect swine, thereby reducing the animals' weight but increasing their food consumption. *Ascaris* may also infect man, causing a chronic infection known as ascariasis. These organisms, *Ascaris* sp., were found in such small numbers that no statement will be made concerning their fate in

EPA-287 (Cin.)
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TYPING GUIDE SHEET

either the anaerobic or aerobic digestion process. Capillaris sp. and Toxocaris leonia were also isolated in very small numbers.

In summary, it can be stated that the aerobic thermophilic digester exhibited far superior performance over the mesophilic anaerobic digester(s) with respect to the inactivation of pathogenic bacteria, viruses, and parasites. The thermophilic aerobic digester yielded complete inactivation of Salmonella sp. and total plaque forming units, i.e., below the limits of detection, on all but one occasion, thereby demonstrating the system's ability to control these pathogens. Parasite numbers were significantly reduced by the aerobic thermophilic system, but not completely. The inability of the system to deliver parasite-free sludge may be indicative of the need for a multi-stage system with heat exchangers to prevent possible short circuiting (hydraulic) and increased reactor temperatures for control of the environmentally resistant parasites, e.g., Ascaris sp.

## HEAVY METAL INTERACTIONS

At the outset of this study, it was felt that the high temperature and the oxygen environment might change the form or even solubilize the heavy metals in the sludge. The screening data indicate that most of the heavy metals remain associated with the particulates during the aerobic thermophilic digestion process. However, preliminary experiments indicate that a further chemical treatment may be capable of efficiently extracting heavy metals from these high temperature treated aerobic sludges. Additional studies are planned for this area.

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## SECTION 9

## ENGINEERING SIGNIFICANCE AND ECONOMIC CONSIDERATIONS

The results of this study have shown, at a large scale, that simple air-aeration system can be designed to achieve high temperatures that result in high rates of stabilization and pasteurization of liquid sewage sludge. Practical questions such as the cost competitiveness, design requirements, and scale-up needs remain to be clarified. Although a complete economic analysis is beyond the scope of this study, considerations of several aspects indicate that this technology has significant potential for providing a reliable and perhaps less costly alternative to existing liquid slurry treatment systems.

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The components of an air aeration system that would be capable of autoheating to thermophilic temperatures with most sewage sludges is shown in Figure 87. The key to successful conservation of energy in the system is the use of the self-aspirating aerators that are capable of high oxygen transfer efficiencies. These kinds of aerators operate without blowers or compressors, and are therefore simple and relatively maintenance-free. Although the DeLaval centri-rator was tested for the majority of the experimental period, the Midland-Frings is shown in Figure 87. It operates close to the bottom of the tank. These two units appeared to achieve about equal oxygen transfer efficiencies, but the Midland-Frings was a much more effective mechanical mixer. It is doubtful as to whether the centri-rator would operate effectively in geometric configurations different from that tested. This would limit the centri-rator's usefulness in large tanks or innovative applications such as insulated lagoons.

Since grit and other materials that might damage the Midland-Frings aerator may settle, it is necessary to set the unit about 0.5 meters above the bottom of the reactor. Also, a way of removing the aerator for maintenance and cleaning should be provided with both units. The influent and effluent pipelines need to be designed to manage the dense foam that will form, e.g., place them below liquid level. A freeboard, one-and-one-half meters deep, should be provided above the liquid surface for foam accumulation. A mechanical foam cutter with a power of 1 to 2 horsepower for every 14 to 19 square meters was capable of controlling foam generated from animal wastes and sewage sludge. It should be emphasized that foam control in this process is significantly different than low temperature aeration systems and must be an integral part of any design.

Efficient insulation of the reactor is also a necessary component of the design. In this study, 5 to 10 cm of sprayed-on polyurethane foam allowed the use of autoheated phenomenon in a cold climate. Feeding of sludge at freezing temperatures with air temperatures far below freezing do not prevent

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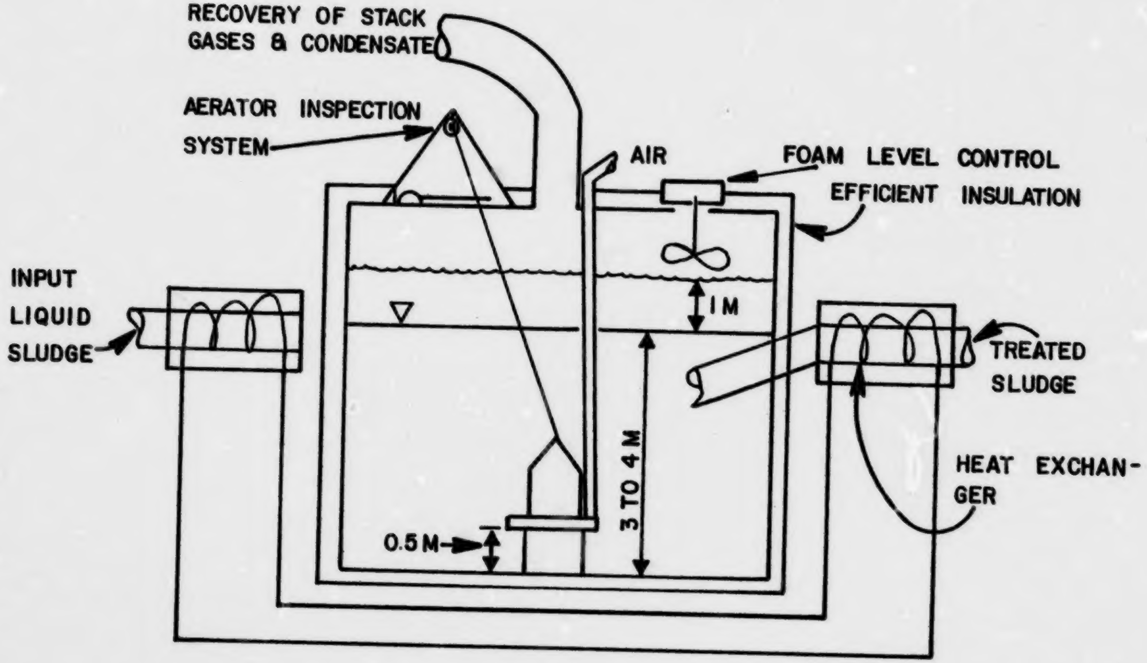
RECOVERY OF STACK GASES & CONDENSATE
AERATOR INSPECTION SYSTEM
AIR
FOAM LEVEL CONTROL
EFFICIENT INSULATION
INPUT LIQUID SLUDGE
1 M
TREATED SLUDGE
3 TO 4 M
HEAT EXCHANGER
0.5 M

Figure 87. Thermophilic aerobic digestion system.

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the use of this technology when adequate methods of energy conservation are applied. The added cost of insulation will be discussed later.

In applications where there is a limited amount of energy available (in other words, where the concentration of the biodegradable organics is low), heat exchangers may be a necessary part of the design. An example may be where large quantities of poorly settling waste activated sludge only thickens to concentrations of around one percent solids. Another instance where a heat exchanger would be useful is in a cold climate where high rate treatment is utilized.

A final component of a practical design would be capture and control of the effluent gases. Although there is a small quantity of aerating gas, it contains ammonia, is saturated with water vapor, and has the humus odor typical of activated sludge. Other volatile pollutants may also be present. Also, start up and shutdown with sludge which arrives in a septic state may initially generate obnoxious odors when autoheated to high temperatures. Simple stacks 3 to 6 meters high will probably allow condensation and elimination of most problems. However, in other cases, it might be necessary to use simple scrubbers on the effluent gas.

Finally, where efficient pathogen kill is required, interruption of the continuous flow from the completely mixed systems is necessary. In all completely mixed systems, a small fraction of the flow escapes in the effluent after only a short time in the reactor. Thus, a unit such as Figure 87 would be operated intermittently to achieve maximum kill, or a series of completely mixed reactors could be utilized to decrease this "short-circuiting" effect.

## DESIGN VOLUME REQUIREMENTS FOR STABILIZED SLUDGE

In order to compare the results of the autoheated digestion system with traditional treatment, it is necessary to develop a common basis for comparison. What is the definition of stabilized sludge? Presently, the most common answer might be about 40 percent total volatile solids destruction. However, it is clear that there are a number of problems with this definition. The biodegradability of sludges produced from different processes varies significantly, as does the sludge from a given facility. This study showed that the biodegradability varied more than 100 percent over the year.

A more qualified question might be, what characteristics should a sludge have under varying conditions of final disposal? In cases where transportation distance is short, odor control is not a problem, and pathogens and other toxics do not present hazards, no treatment of the raw sludge may be necessary. Direct soil injection of raw or unstabilized sludge is practiced in some instances such as the above.

In general, it would appear that there are three separate requirements based primarily upon the final disposal consideration for differing degrees of sludge stabilization: (1) no treatment after separation from the wastewater, (2) minimal treatment for odor and hazard control, and (3) maximum

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treatment to avoid the potential of creating a nuisance or hazard in more risk-prone circumstances. Practical interpretations of the degree of stabilization required to meet the second condition could probably require, as a minimum, odor control and pasteurization of the sludge. Maximum treatment would probably require odor control, pasteurization, and some level of biodegradable organic matter destruction yet to be defined. Although the criteria to define these two conditions were not developed here, a qualitative design approach could be outlined. In all cases tested, the highest loading rates (at 2.9 day HRT) resulted in odor destruction and achievement of pasteurization temperatures. The longer term studies defined the kinetics, so that the degree of organic matter oxidation could be estimated.

Another factor needs to be considered in design volume requirements: the effect of mixing on stabilization efficiency in completely mixed reactors. The simple first order decay model used to describe the decay of biodegradable organics in a completely mixed tank is as follows:

$$\frac{S_e}{S_o} = \frac{1}{1 + Kt} \qquad (1)$$

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where $S_e$ = biodegradable organics in the effluent, mass/volume
$S_o$ = biodegradable organics in the influent, mass/volume
$K$ = reaction rate coefficient, time$^{-1}$
$t$ = (hydraulic) retention time, time

Several completely mixed reactors in series of equal volume are described by:

$$\frac{S_e}{S_o} = \left[\frac{1}{1 + K\left(\frac{t}{n}\right)}\right]^n \qquad (39)$$

where $n$ = number of completely mixed reactors.

An infinite series of completely mixed units approaches a plug flow reactor which is equivalent to a batch reactor. In some cases, it may be desirable to use several batch reactors. Effective control of batch systems composed of several reactors is a practical alternative today because of the control made possible through the application of microprocessors. The decay in a plug flow or batch system is described by the exponential relationship:

$$\frac{S_e}{S_o} = e^{-Kt} \qquad (40)$$

Consideration of the influence of multi-staged completely mixed reactors indicates that they are much more efficient than one large completely mixed unit. Figures 88 and 89 illustrate the difference between one stage and multiple series reactors. In Figure 88, each series reactor has a one-day

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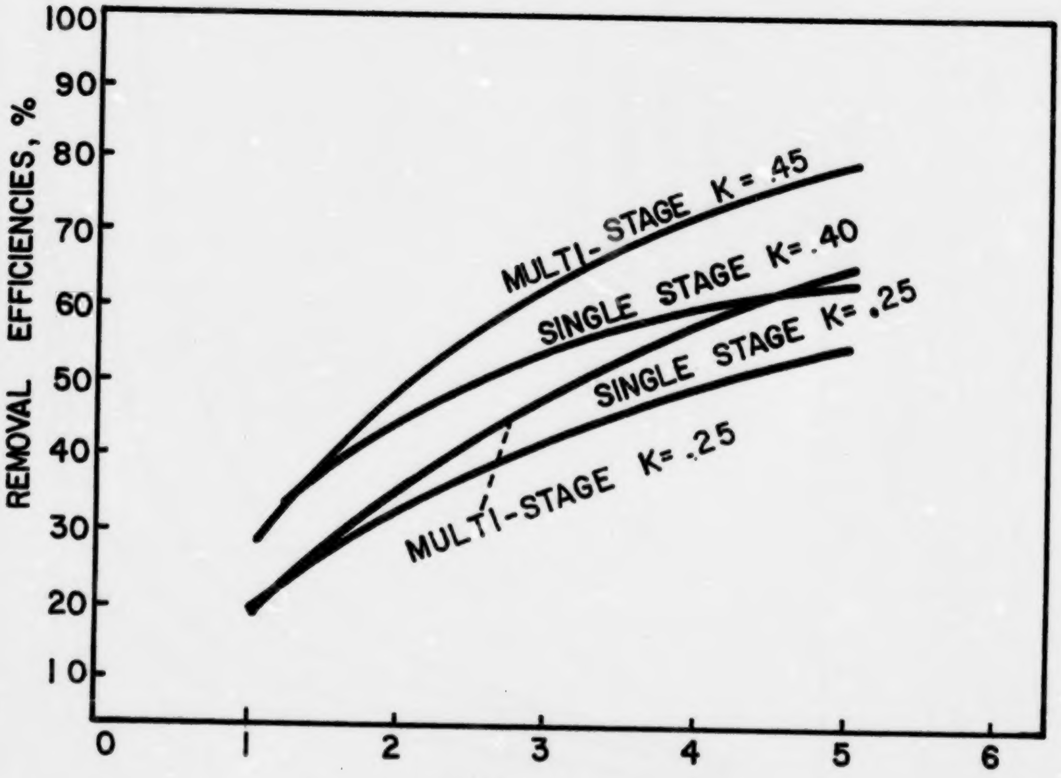

Figure 88. Influence of single and series multiple stage, completely-mixed reactors on substrate removal efficiencies with two different substrate removal rate coefficients. Each series reactor has a one day HRT.

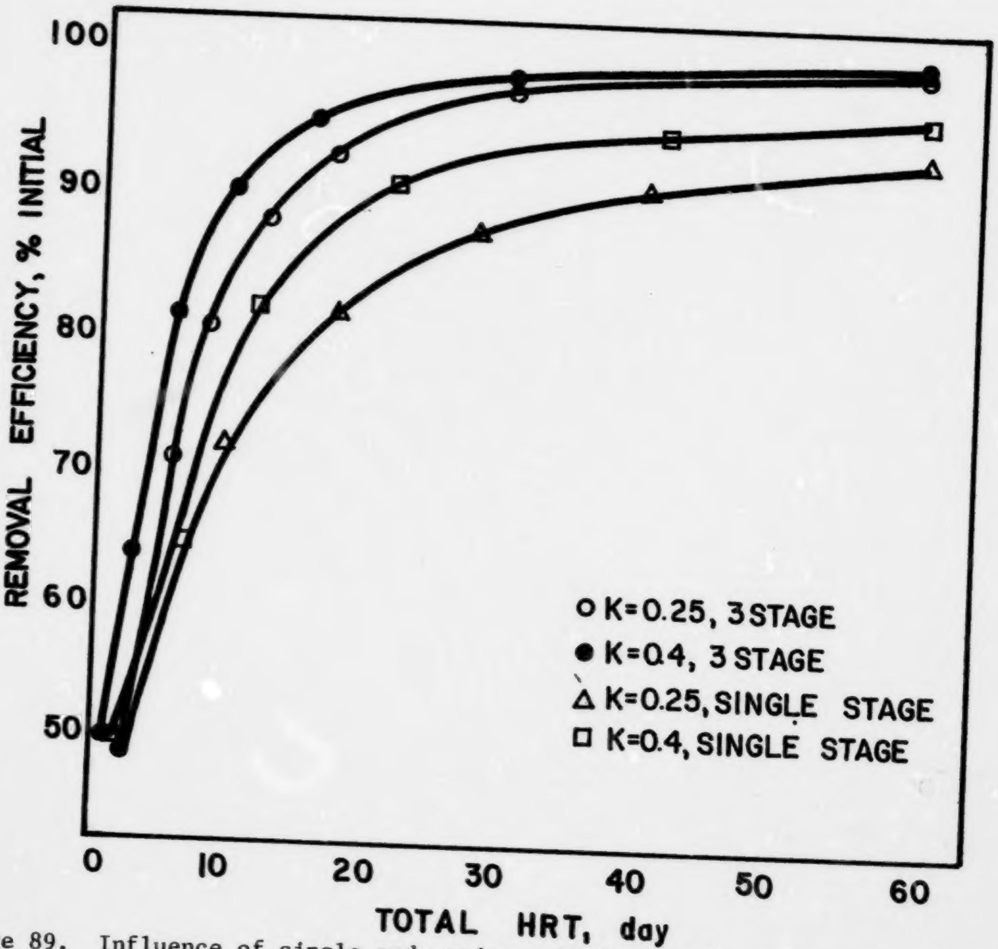

Figure 89. Influence of single and series multiple stage, completely-mixed reactors on substrate removal efficiency with two different substrate removal rate coefficients. Each series reactor system has 3 stages with one-third the volume (per stage) of a one-stage system.

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retention time so that a single stage 5-day unit is compared to a 5-stage series reactor, each with a one-day retention time. Figure 89 differs, in that the single stage reactor is compared to the results obtained with a 3-stage system. For example, the results obtained at a 15-day retention time are compared to those obtained from a 3-stage reactor, each with a 5-day retention time. Note that all differences in removal efficiency are obtained from the mixing effects only.

In the well mixed runs with the Midland-Frings reactor, where the average dissolved oxygen concentrations were higher, the biodegradable substrate removal rate (K in Eq. (1)) reached a value as high as 0.8 $\text{day}^{-1}$. But in the laboratory units, the organic matter destruction rate coefficient was often 0.2 $\text{day}^{-1}$ to 0.25 $\text{day}^{-1}$. The calculated removal characteristics at low removal rates ($K = 0.25\ \text{day}^{-1}$) are compared to those attained at a higher value ($K = 0.4\ \text{day}^{-1}$) in Figures 88 and 89. The data in Figure 89 indicate that there is little difference between the reactors when the removal efficiency is assumed to be 50 percent. However, the higher the required efficiency of substrate removal, the more significant become the results obtained from a series reactor system. If 80 percent biodegradable organic removal is required, a three-stage reactor, each having a 1.5-day HRT (total HRT of 4.5 days) would achieve the same results as a one-single reactor with a retention time of 12 days. Two modifications to the system would be suggested if 1.0 to 1.5 day retention period reactors were to be used. First, in order to provide a viable and significant amount of thermophilic microorganisms in the first reactor, recycle from the third stage back to the influent is suggested. Second, in order to minimize the temperature shock and to provide more reliable temperature control, a heat exchanger designed to preheat the influent should be utilized. In some instances (especially where the sludge is highly concentrated), it may be necessary to cool the reactor to prevent temperatures from rising to inhibitory levels (greater than 70°C).

The most significant difference in substrate removal caused by mixing is created when maximum efficiencies are required. For example, if 95 percent organic matter destruction were required, the total system retention time required would be as follows in days (as shown in Figure 90):

| | Single Reactor | Three-Stage Reactor |
|---|---|---|
| $K = 0.25\ \text{day}^{-1}$ | 70 | 22 |
| $K = 0.40\ \text{day}^{-1}$ | 46 | 12 |

Thus, a system composed of three 4-day HRT reactors in series would result in the same organic matter destruction efficiencies as a single unit with an HRT of 46 days when the biodegradable substrate removal rate is 0.4 $\text{day}^{-1}$. It also should be noted that the optimum series reactor systems is composed of reactors of equal volume.

There are several practical constraints on the application of this kind of analysis. Small completely mixed series reactors are more susceptible to toxic effects and are little used in waste treatment. Also, it is assumed that organic removal rate coefficients remain constant and this may not be valid.

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EPA-287 (Cin.)
(4-76)

## ECONOMIC COMPARISON TO CONVENTIONAL TREATMENT

Although mesophilic anaerobic digestion and ambient temperature aerobic digestion may not achieve the same results, their costs will be included for comparison purposes. A review of some cost data for sludge treatment is given in Table 33. It would appear that aerobic digestion has a cost advantage at low flows, and that anaerobic digestion is more cost effective at high flow rates. The complexities introduced by pure oxygen utilization may not be warranted in view of the results obtained in this study with air aeration, except in cases where oxygen is already used by the plant.

Additional comparison of the cost comparison in terms of capital, operation and maintenance, and energy consumed is shown in Figure 90. The most obvious difference between the two conventional digestion requirements is the energy cost for aeration. At $0.03 per kW-hr, the cost of energy for aeration is about $10.00 per ton of sludge or $0.25 to $0.35 per person per year. The energy cost of using pure oxygen is about the same. In aerobic digestion, where nitrification occurs, the aeration energy costs may be more than double that shown. Since nitrification cannot occur at thermophilic conditions, additional cost of aeration for this purpose is not a concern here.

If a cost comparison were to be made on the basis of a system that achieved the accepted value of 40 percent TVS destruction, the data in Figure 89 and Table 33 could be used to develop comparison information. If a sludge has a maximum biodegradable fraction of 60 percent of the TVS, 40 percent destruction would mean that two-thirds of the maximum biodegradable mass would be oxidized. A three stage system operating at 1.3-day HRT in each stage would be able to achieve this removal efficiency, as estimated in Figure 89. This would result in a considerable decrease in capital cost over other systems with detention times varying from 12 to 20 days. At low sewage flow rates, the savings that could be attributed to building reactor volumes one-half to one-quarter that required by other technologies would result in a savings of $10.00 to $20.00 per ton of sludge processed as a result of decreasing the capital investment for the reactor.

The major difference between conventional aerobic digestion at low temperatures is the efficient insulation and covered reactor required for this system. However, insulation is not a major cost component in construction and implementation of large biological reactors. This can be illustrated by comparing the added cost of insulation to the increased cost per ton of sludge processed. An insulation value capable of efficiently conserving the reactor energy ($R \geq 25$) can be obtained for less than $60 per $m^2$ ($2/$ft^2$) of reactor surface. For a sewage flow value of about $0.18 m^3$/sec (4 MGD), insulation costs would increase the sludge treatment cost by about $0.91 per Mg ($1.00 per ton). This is the primary cost difference between a system using conventional aerobic digestion and the requirements for an autoheated reactor system. An insulated reactor top is also needed with the autoheated system.

## TABLE 33. COMPARISON OF DESIGN SPECIFICATIONS AND REPORTED COST FOR SEWAGE SLUDGE TREATMENT.

| | | PROCESS DESIGN CRITERIA | | | | | | | COSTS, $/dry ton processed | | | | |
|---|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Sludge Treatment Process | Equivalent Wastewater Flow Rate (MGD) | Solids Destruction (% VSS) | HRT (Day) | Loading Rate VSS (lb/ft$^3$-day) | kg/m$^3$-day | Reactor Vol. Required (ft$^3$/cap) | m$^3$/cap | Temp (°C) | Capital | Operation and Maintenance | Energy | Total | References |
| Aerobic Digestion | 1 | 40 | 18-20 | 0.02-0.4 | - - | 3-4 | - - | 20 | 89 | 3.25 | 15 | 107 | U.S. EPA 1974, 1976, 1977 |
| | 10 | 40 | 18-20 | 0.02-0.4 | | 3-4 | | | 37 | 1.17 | 15 | 53 | |
| Aerobic Digestion Thermophilic with Oxygen | 1.4 | 40 | 10 | 0.3-0.5 | - - | - - | - - | >40 | 31 | 17.6 | 15 | 64.6 | American Society of Civil Eng. 1974 |
| | 14 | 40 | 10 | 0.3-0.5 | | | | >40 | 5.2 | 2.7 | 15 | 22.9 | |
| Anaerobic Digestion Two Stage | 1 | 35-50 | 20 | 0.04-0.4 | - - | 2.6-6 | - - | 35 | 148 | 50 | 0.6 | 199 | U.S. EPA 1974, 1976 1977 |
| | 10 | 35-50 | 20 | 0.04-0.4 | | 2.6-6 | | 35 | 44 | 9.8 | 0.27 | 54 | |
| | 100 | 35-50 | 20 | 0.04-0.4 | | 2.6-6 | | 35 | 18 | 5.86 | 0.39 | 24 | |
| Anaerobic Digestion Two Stage Thermophilic | 1 | 60-70 | 15 | 0.1-0.2 | - - | - - | - - | 50 | 240 | 50 | 2.4 | 292 | As above + U.S. EPA 1975 |
| | 10 | | | | | | | | 59 | 14 | 2.4 | 75 | |
| | 100 | | | | | | | | 31 | 3.9 | 2.4 | 37 | |
| Anaerobic Digestion Two Stage | 1 | - - | - - | - - | - - | - - | - - | - - | 35 | 25 | - - | 60 | Van Note et al. 1975 |
| | 10 | | | | | | | | 10 | 4 | - - | 14 | |

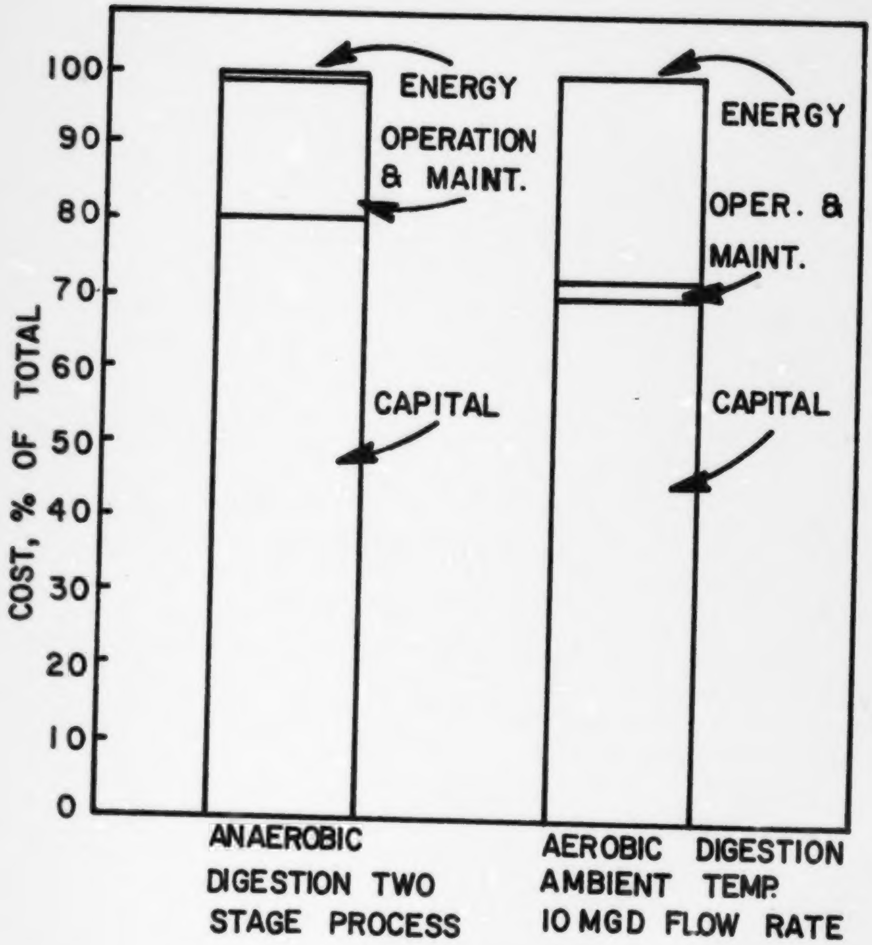

Figure 90. Comparison of capital, operation, and maintenance cost(s) for sludge treatment.

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## SUMMARY

The engineering significance indicates that an alternative of utilizing a simple autoheated aerobic digestion system is a viable alternative to ambient temperature aerobic digestion and anaerobic digestion systems; it has the potential of improved treatment and pathogen control at a cost less than existing approaches.

Based on this preliminary economic analysis, it would appear that the following conditions would be met with this system:

-autoheating would occur with hydraulic retention times as low as 2.5 days with primary and waste activated sludge containing solids at volatile solids concentration greater than 20 gm/ℓ with a biodegradable fraction greater than 50 percent of the TVS;

-TVS destruction efficiencies should be achieved in less than a 5-day HRT by the autoheated air system that are now achieved by conventional aerobic and anaerobic treatment systems, if the temperature and sludge characteristics meet the above characteristics;

-if total hydraulic retention periods equal to 15 or 20 days are utilized in a three stage liquid reactor, destruction of the biodegradable solids should be complete, or about 50 to 70 percent of the TVS should be oxidized as opposed to a 40 percent destruction efficiency at comparable detention times in mesophilic anaerobic and aerobic treatment processes;

-the economics would appear to favor this system. When comparable TVS destruction is required, and when this is greater than 80 percent of the biodegradable fraction, a volume reduction of over 50 percent would result. Loading rates for this system would vary between 6-13 kg/m$^3$-day (0.4 and 0.8 lb TVS per cubic foot of reactor per day). The savings in reactor volume reduction would exceed \$10.00 per ton of sludge processed. Added insulation cost is the major difference between conventional aerobic digestion and the system introduced here. This adds about \$1.00 per dry ton to the total processing cost. Finally, since nitrification cannot occur at thermophilic temperatures, aeration energy cost savings over a nitrifying ambient temperature aerobic system would be about \$10.00 per dry ton;

-treatment for odor control and efficient pathogen destruction only, without regard to dewatering or dewatering influences, should be achieveable in a three stage system with a total retention time only slightly exceeding 3 days, where about 10 percent flow is recycled, and a heat exchanger is utilized;

-because of the simplicity of the self-aspiration air aeration system, there appears to be little benefit in using pure oxygen, except in cases where the organic concentration is exceptionally dilute (less than 1%) or in a pure oxygen activated sludge plant.

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# APPENDIX A

## PATHOGEN ISOLATION AND DETECTION METHODOLOGY

A synopsis of the various analytical methods used in this study for the isolation and identification of pathogenic microorganisms is presented in this appendix. The procedures presented here were provided for this report by the various laboratories and principal investigators performing the analyses.

### PROCEDURE FOR THE DETECTION OF PATHOGENIC BACTERIA AS UTILIZED BY CORNELL UNIVERSITY VETERINARY DIAGNOSTIC LABORATORY (Shin, 1978)

I. Isolation of Pathogens

- A. Selective growth on solid media
  - 1. Differential media: MacConkey's agar
  - 2. Selective media
    - (a) Brilliant Green agar
    - (b) Xylose lysine desoxycholate agar
- B. Enrichment broth
  - 1. Dulcitol selenite broth (subcultured to BG agar)
  - 2. GN broth (subcultured to XLD agar)

II. Identification of Pathogens

- A. Colony morphology on agar plate
- B. Biochemical tests
  - 1. TSI slant
  - 2. Enterotubes

III. Enumeration of Pathogens

- A. Sample is diluted with sterile $H_2O$ to $10^{-3}$ and $10^{-4}$ dilutions
- B. 1.0 ml of dilution is inoculated onto selective media
- C. Colonies of confirmed pathogens are counted.

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## PROCEDURES FOR PARASITOLOGIC EXAMINATION OF WASTE SLUDGES AS DEVELOPED BY TULANE UNIVERSITY SCHOOL OF PUBLIC HEALTH AND TROPICAL MEDICINE (Little, 1978)

### Reagents

1. 1% of soln (v/v) of "7X," an anionic detergent
2. $ZnSO_4$ soln; specific gravity 1.20, adjusted with hydrometer
3. Alcohol - Acid Soln; 35% Ethyl Alcohol +0.27% (v/v) $H_2SO_4$
4. Ether - Anhydrous
5. Siliclad (Clay Adams), for coating glassware with silicone.

### Materials

1. Beakers, 1000 ml, tall form, Berselius, Pyrex
2. Beakers, 1000 ml, low form, Kimax
3. Sieves, 7 in., gauges 10, 20, 50, 100, and 150
4. Funnel, powder filling, 150 mm, Nalgene
5. Applicator sticks, wooden
6. Centrifuge tubes, conical, polypropylene, 50 ml
7. Centrifuge tubes, conical, "Pyrex," 15 ml
8. Pipettes, Pasteur, disposable, 5-3/4 in.
9. Microscope slides, 3 x 2 in.
10. Cover glasses, 22 x 30, 24 x 40, 24 x 50 mm
11. Wash bottles, plastic, 500 ml and 1000 ml
12. Aspirator carboys, 8 l, Nalgene
13. Aspirator, bottle, 4 l, Nalgene.

### Equipment

1. Centrifuge, IEC, with head and cups for 50 ml tubes
2. Centrifuge, Clinical model, with head for 15 ml conical tubes
3. Magnetic Stirrer
4. Ultrasonic cleaner, 11-1/2" x 5-1/2" x 6"
5. Microscope, Compound, A.O.
6. Balance, Double Beam, Harvard.

### Procedure

#### A. Thickened Sludges

1. Place measured volume of sample (50 to 100 ml depending upon amount of solids) in 1000 ml beaker, low form.

2. Add about 250-300 ml of 1% soln of "7X" and mix for at least 5 min. in magnetic stirrer.

3. Strain homogenized sample through the following series of sieves: 10, 20, 50, 100, and 150. Sieves placed on funnel over another 1000 ml beaker and sample washed through each sieve with aid of stream from a wash bottle containing 1% "7X."

4. After sample has been passed through each sieve, it is placed in a 1000 ml beaker, tall form, and allowed to settle for at least one hour.

5. Supernatant is decanted and discarded and sediment is placed in 1 (more, if necessary) 50 ml centrifuge tube.

6. Sample is centrifuged at 2000 rpm for at least 2 min. and supernatant poured off.

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7. If packed volume of sediment is 7-8 ml or less, proceed as below; if volume is more than this, the sediment should be resuspended in tap water evenly divided between 2 tubes, and centrifuged as above.

8. Add $ZnSO_4$ soln to packed sediment, mix thoroughly with applicator sticks, increase volume to 50 ml and centrifuge at 2000 rpm for 3 min.

9. After centrifuge has come to a full stop and soln in tubes has stopped swirling, pour top 30 to 40 ml of supernatant into a 400 ml beaker. During pouring off of supernatant, rotate tube in order to allow all of material floating on surface to be poured off with supernatant.

10. Dilute poured off supernatant to about 250 ml with tap water.

11. Place diluted supernatant in 6 or more 50 ml tubes and centrifuge at 200

12. Discard supernatant by pouring off and transfer sediment to one or more 15 ml centrifuge tubes, using stream of tap water from a wash bottle to completely wash all of sediment from each tube. Fill to near top with tap water.

13. Centrifuge at full speed in Clinical Centrifuge for 2 min., discard supernatant. If more than one tube is used, combine sediment into one tube unless combined sediment would exceed 1.5 ml.

14. Add 10 ml of Alcohol-Acid soln to sediment (packed sediment should not exceed 1.5 ml; if so, divide sediment into 2 or more tubes), and mix thoroughly with applicator stick.

15. Immediately add 3 ml of ether, place rubber stopper in tube, and mix by shaking vigorously for 15 sec.

16. Remove stopper and centrifuge for 2 min. at full speed.

17. Use applicator stick to ring tube to free plug at ether Alcohol-Acid interface, and pour off entire supernatant.

18. Transfer sediment to one or more slides, cover with appropriate size of cover glass and examine under at least 100 x magnification of compound microscope.

19. Count each kind of helminth egg and note whether eggs appear to be viable or whether they appear to be dead (non-viable). Note presence of protozoan cysts, if any, and indicate kind.

20. For reporting, convert counts to number of organisms per gram or kilogram of dry weight.

## GENERAL COMMENTS

After each use, sieves are washed thoroughly and placed in an ultrasonic cleaner containing appropriate detergent for at least 10 minutes.

In sediments from some anaerobic sludges, the eggs of helminths are darkly stained and difficult to identify when examined microscopically. In such cases, they are bleached slightly by treating with a 2% dilution of commercial bleach (Clorox) for a few minutes, then washed again in water before examination.

Glassware should be coated with Siliclad at appropriate intervals.

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# APPENDIX B

## DETERMINATION OF OXYGEN TRANSFER COEFFICIENT

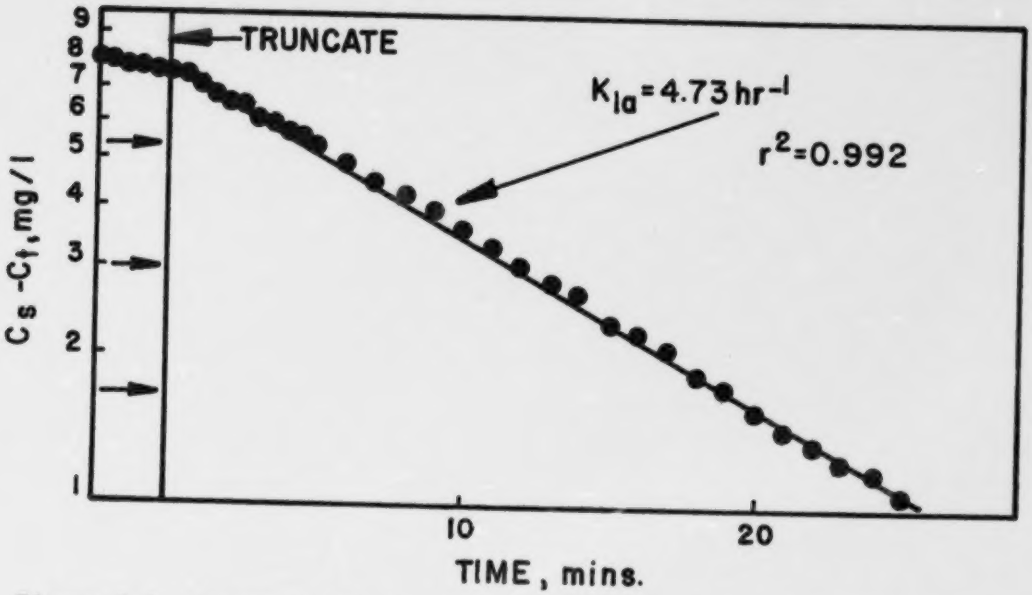

Figure B.1. Determination of $K_{LA}$ at temperature of 24°C using a YSI D.O. meter. Run No. 1.

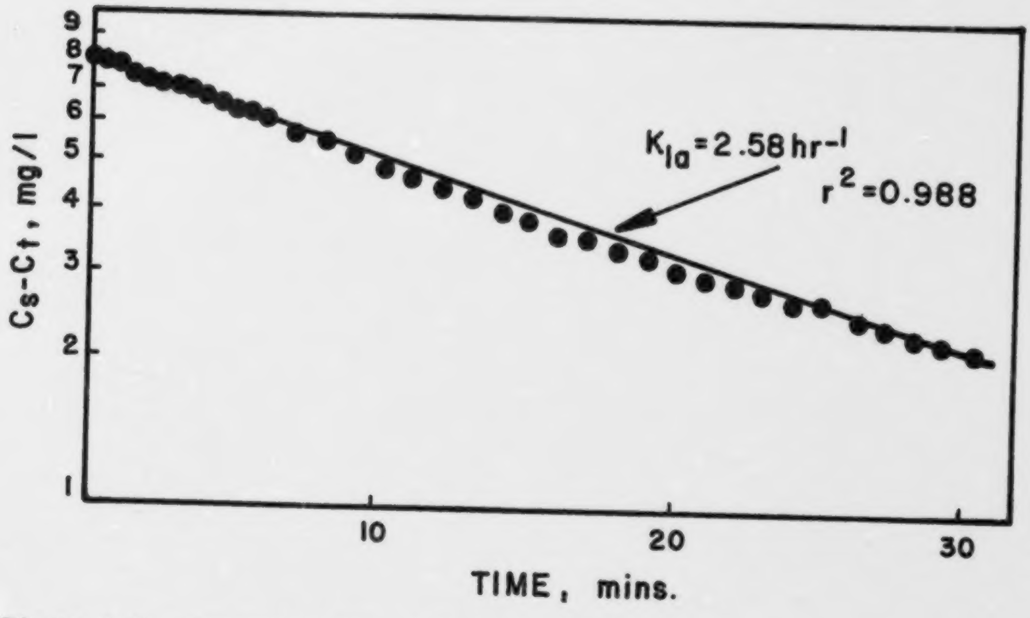

Figure B.2. Determination of $K_{LA}$ at a temperature of 24 °C using a YSI D.O. meter. Run No. 2.

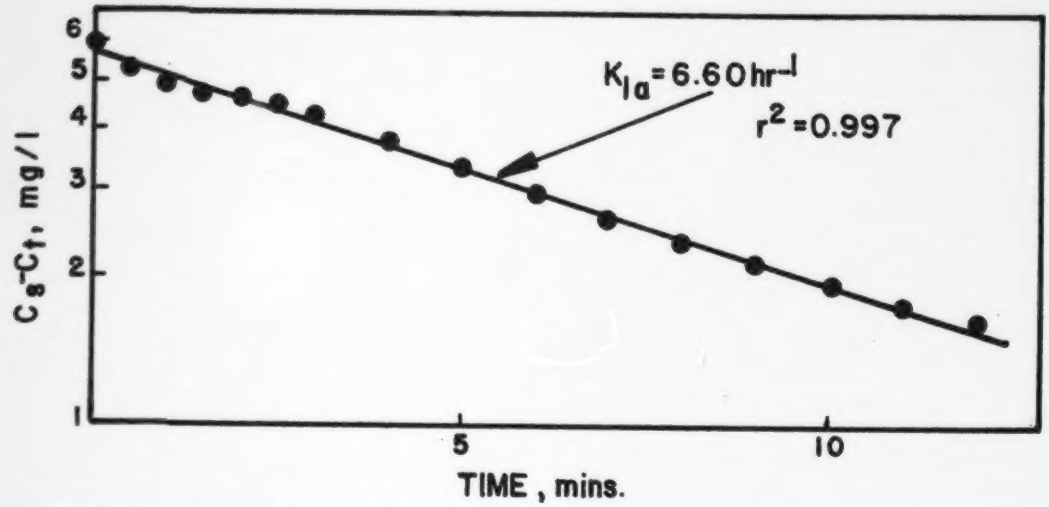

Figure B.3. Determination of $K_{LA}$ at a temperature of 45 °C using a YSI D.O. meter. Run No. 1.

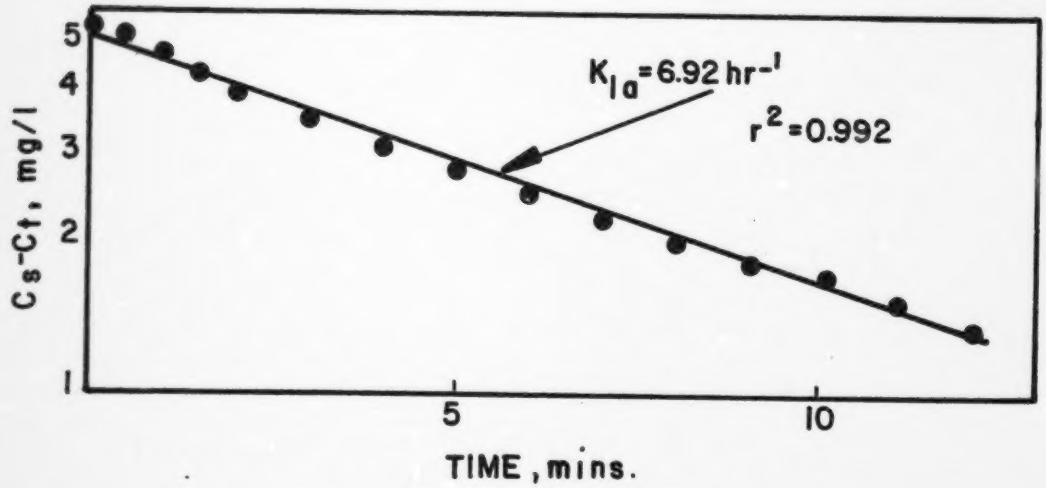

Figure B.4. Determination of $K_{LA}$ at a temperature of 45 °C using an orbisphere D.O. meter. Run No. 1.

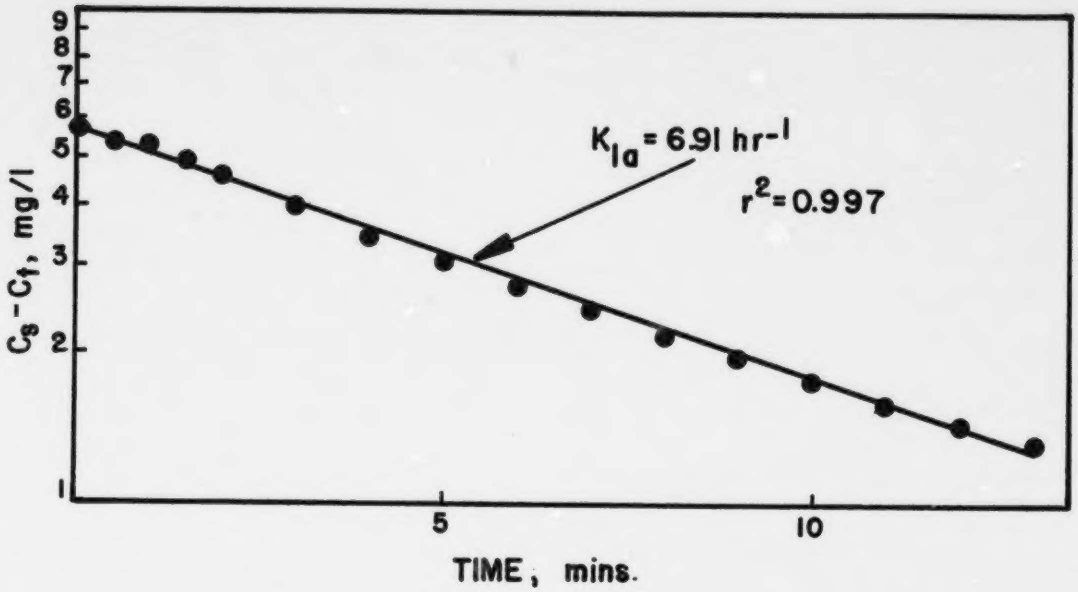

Figure B.5. Determination of $K_{LA}$ at a temperature of 45 °C using a YSI D.O. meter. Run No. 2.

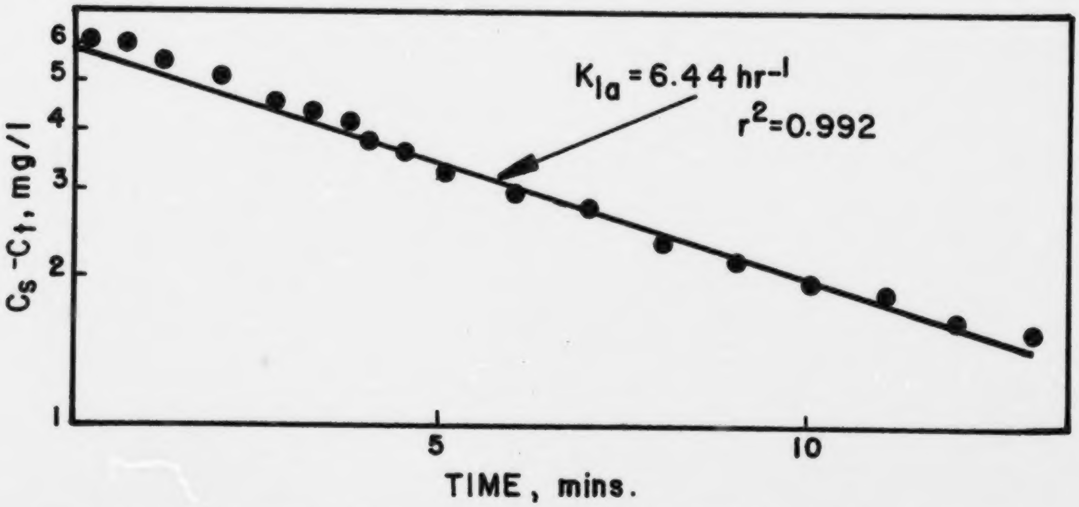

Figure B.6. Determination of $K_{LA}$ at a temperature of 45 °C using an orbisphere D.O. meter. Run No. 2.

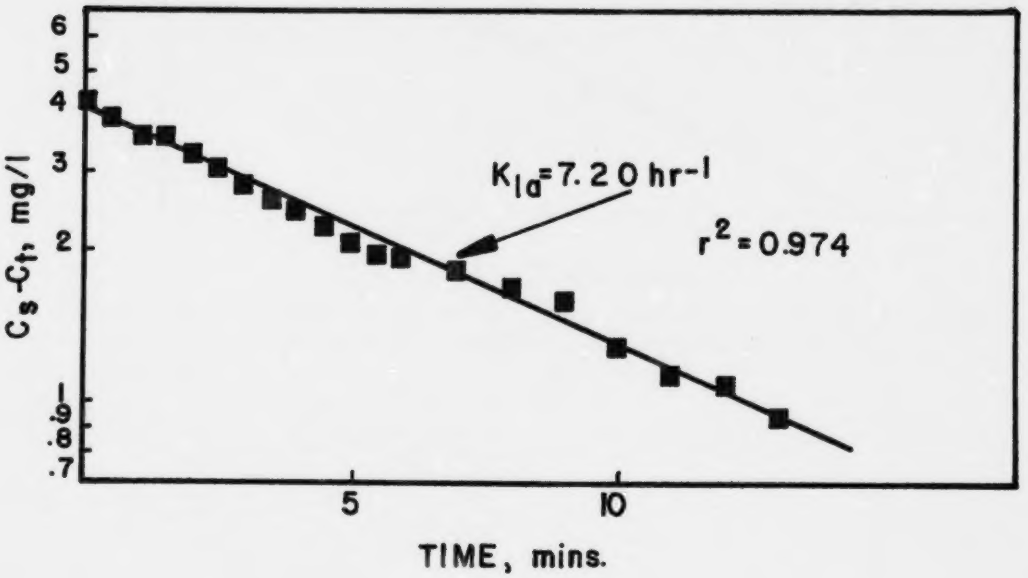

Figure B.7. Determination of $K_{LA}$ at a temperature of 61 °C using an orbisphere D.O. meter. Reaeration Run No. 1.

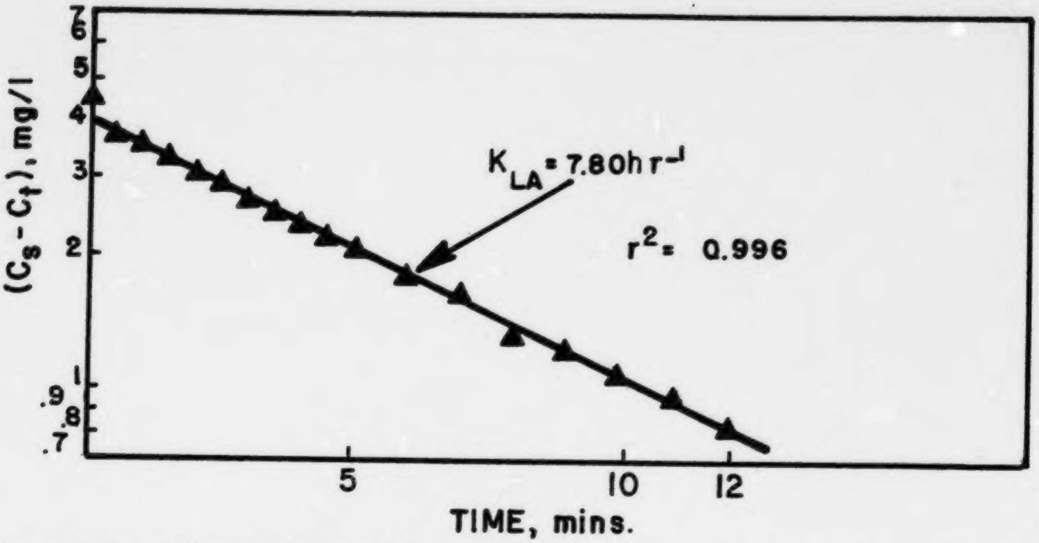

Figure B.8. Determination of $K_{LA}$ at a temperature of 60 °C using an orbisphere D.O. meter. Reaeration Run No. 2.

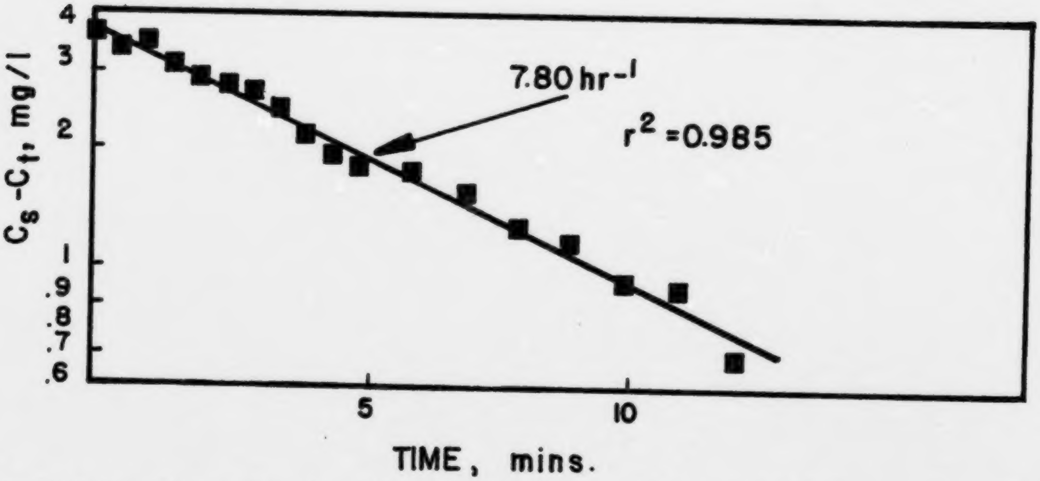

Figure B.9. Determination of $K_{LA}$ at a temperature of 60 °C using the azide modification of the Winkler method for the determination of D.O. concentrations.

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## APPENDIX C

## RESULTS OF BIODEGRADABILITY STUDIES
(Nos. 1-7)

TABLE C-1. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 1, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 36.597 | 34.119 | 21.724 | -- | -- | -- | 63.7 |
| 1 | 28.381 | 26.365 | 17.221 | 22.4 | 22.7 | 20.7 | 65.3 |
| 2 | 25.295 | 24.291 | 14.079 | 30.9 | 28.8 | 35.2 | 58.0 |
| 9 | 16.086 | 14.193 | 9.083 | 56.1 | 58.4 | 58.2 | 64.0 |
| 16 | 11.309 | 14.242 | 7.902 | 69.1 | 58.3 | 63.6 | 55.5 |
| 23 | 7.146 | 12.367 | 6.047 | 80.5 | 63.8 | 72.2 | 48.9 |
| 30 | 5.096 | 13.868 | 6.779 | 86.1 | 59.4 | 68.8 | 48.9 |
| 34 | 5.298 | 11.245 | 5.170 | 85.5 | 67.0 | 76.2 | 46.0 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 52.2 | 50-61 |
| D.O. mg/ℓ | 4.3 | 0.5-5.3 |
| pH | 7.0 | 5.7-7.8 |
| Evaporation, mℓ/day | 300 | 0-600 |
| Airflow rate, ℓ/sec | 0.08 | - |

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TABLE C-2. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 1, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 23.437 | 21.340 | 13.506 | -- | -- | -- | 63.3 |
| 1 | 20.904 | 19.810 | 12.819 | 10.8 | 7.2 | 5.1 | 64.7 |
| 2 | 19.086 | 16.141 | 10.953 | 18.6 | 24.4 | 18.9 | 67.9 |
| 9 | 12.581 | 11.729 | 7.260 | 46.3 | 45.0 | 46.2 | 61.9 |
| 16 | 7.594 | 9.305 | 5.444 | 67.6 | 56.4 | 59.7 | 58.5 |
| 23 | 7.119 | 8.824 | 4.596 | 69.6 | 58.7 | 66.0 | 52.1 |
| 30 | 4.346 | 9.162 | 4.768 | 81.5 | 57.1 | 64.7 | 52.0 |
| 34 | 2.845 | 8.358 | 4.366 | 87.9 | 60.8 | 67.7 | 52.2 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 51.7 | 42-60 |
| D.O., mg/ℓ | 4.1 | 0.1-5.8 |
| pH | 7.1 | 5.7-7.8 |
| Evaporation, mℓ/day | 280 | 0-1100 |
| Air flow rate, ℓ/sec | 0.08 | - |

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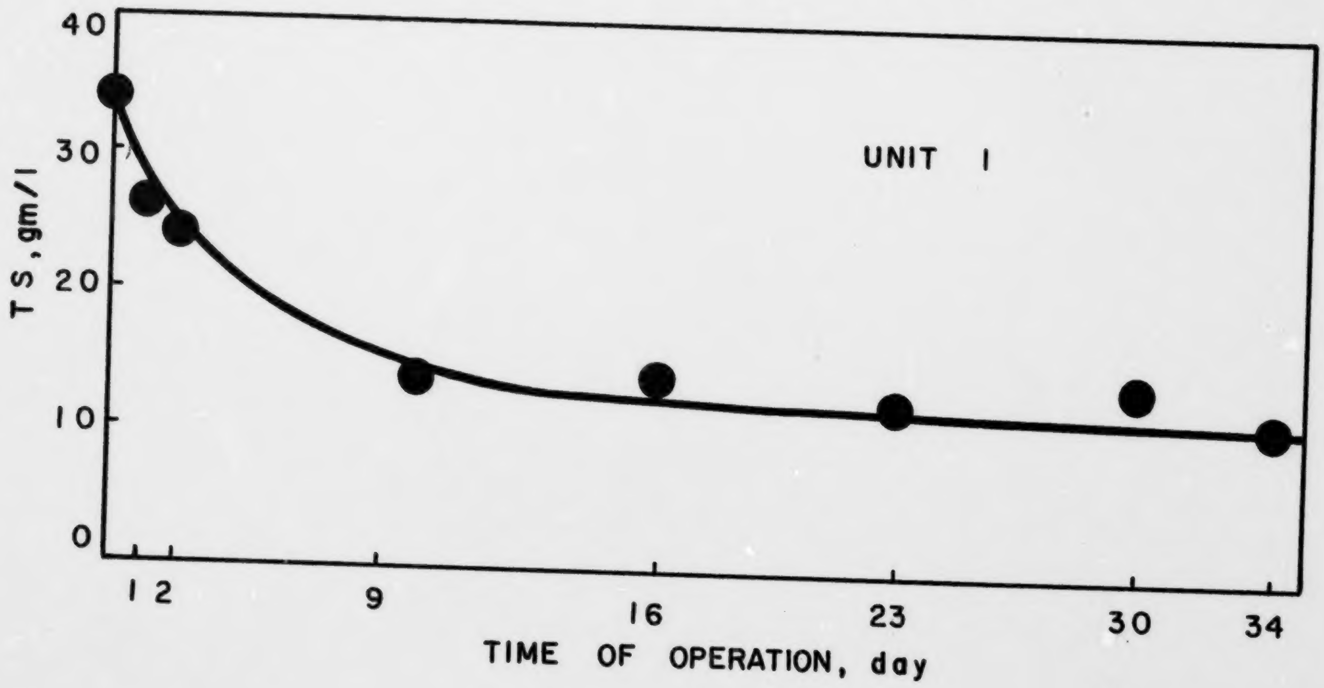

Figure C-1. Total solids (TS) removal as related to time of operation, biodegradability study no. 1.

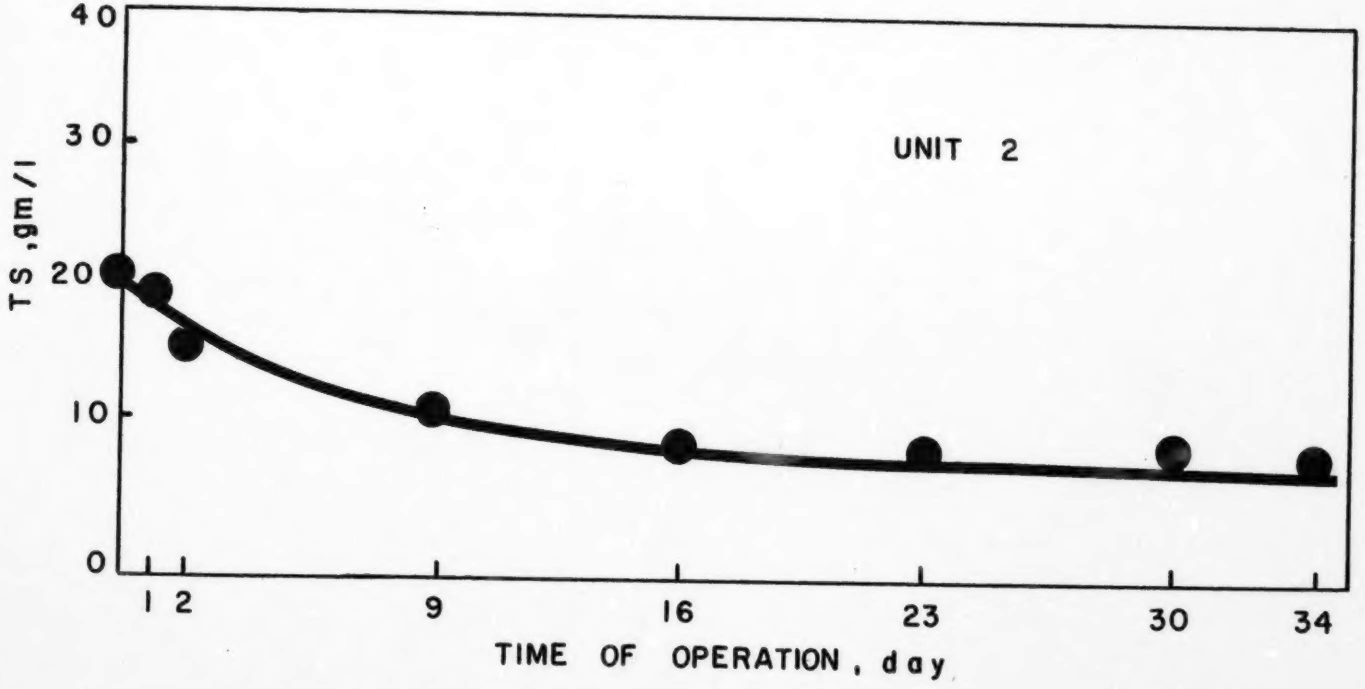

Figure C-2. Total solids (TS) removal as related to time of operation, biodegradability study no. 1.

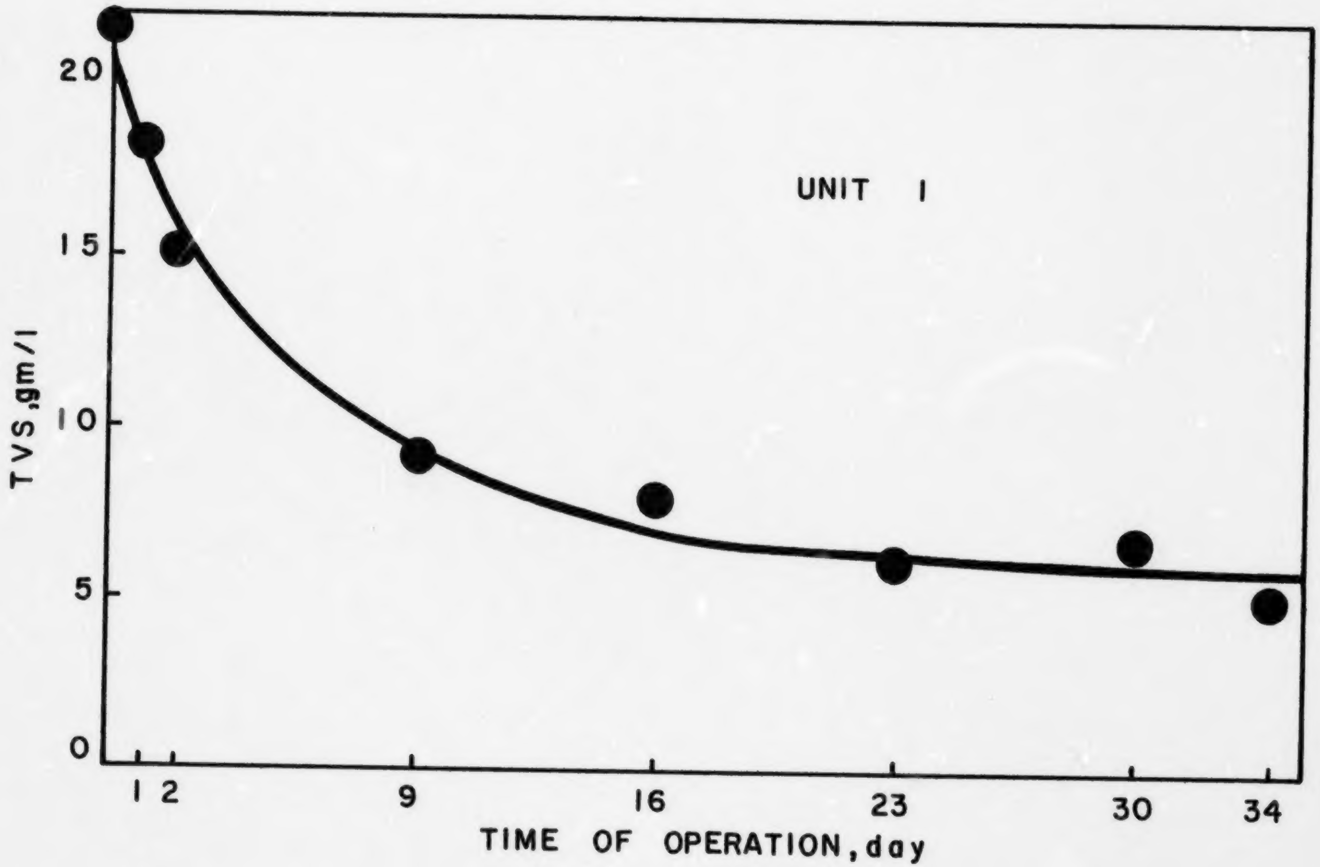

Figure C-3. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 1.

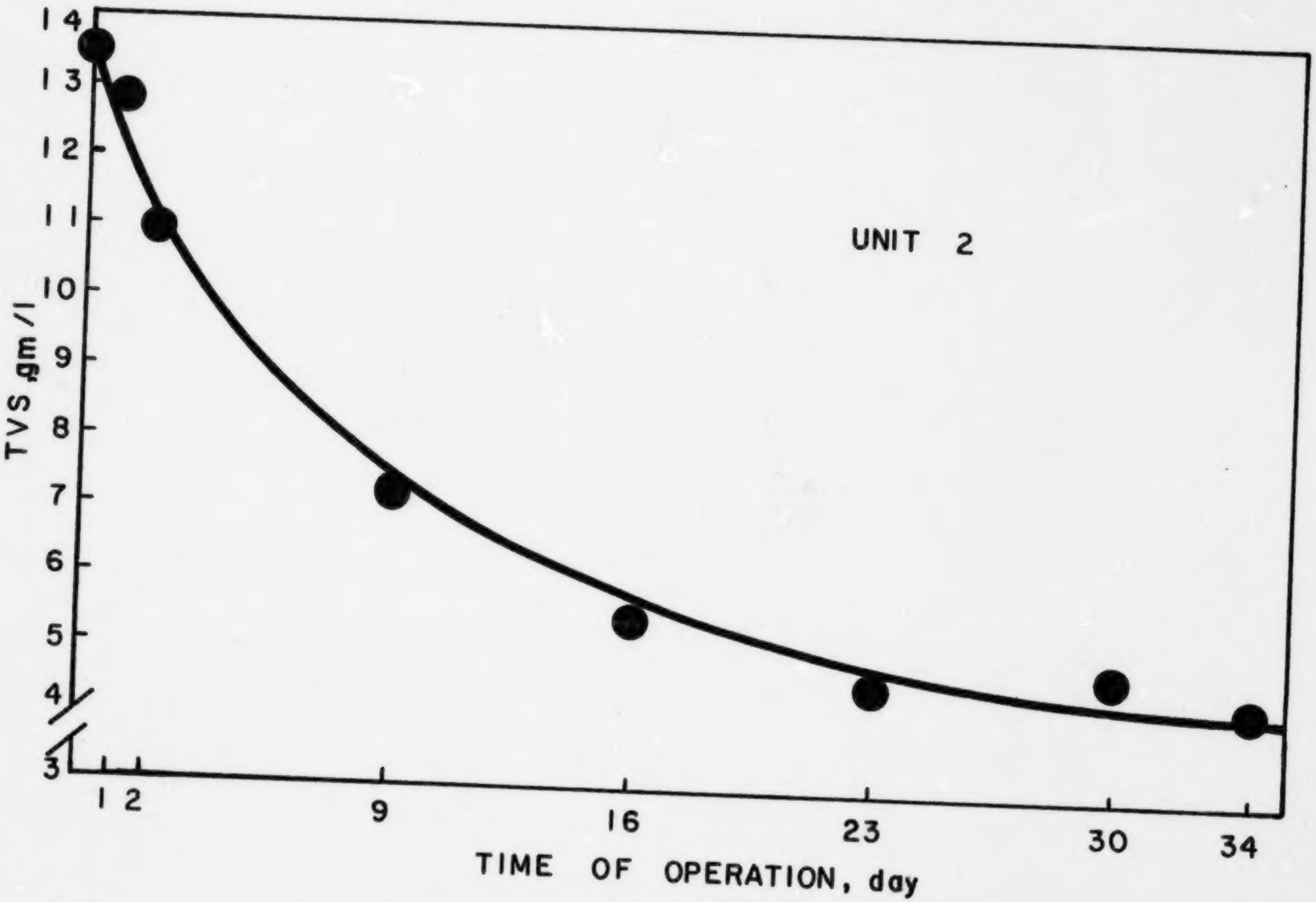

Figure C-4. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 1.

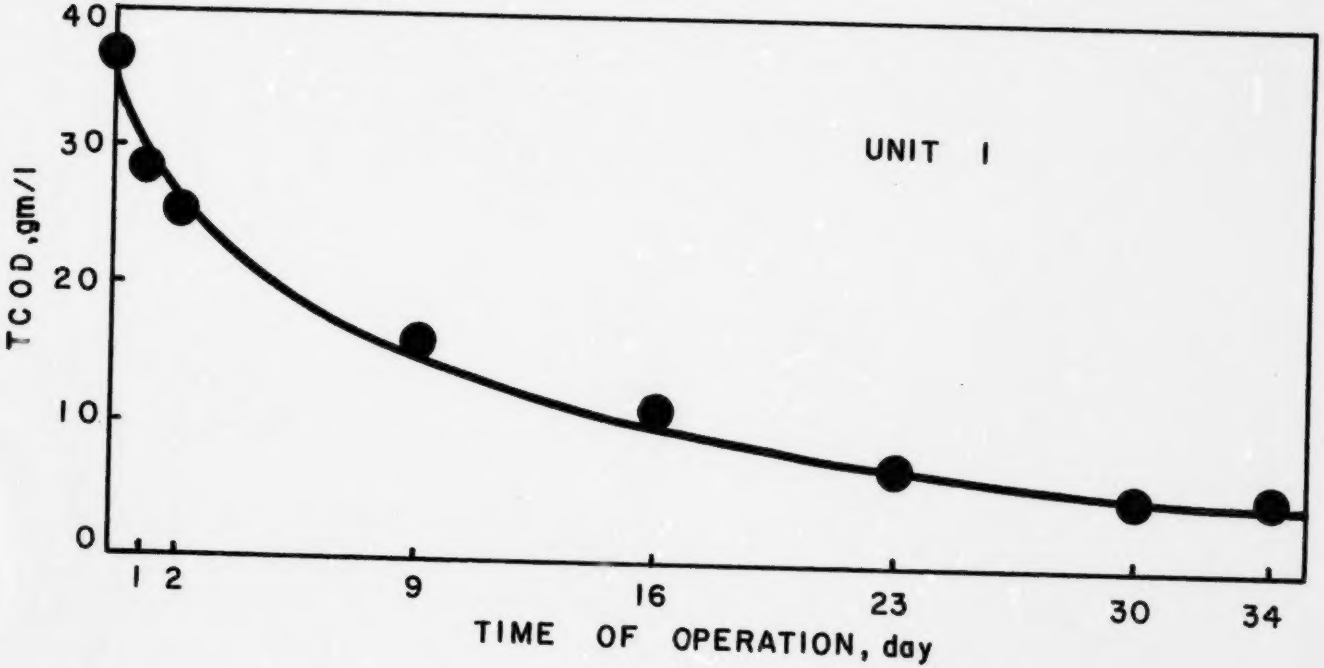

Figure C-5. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 1.

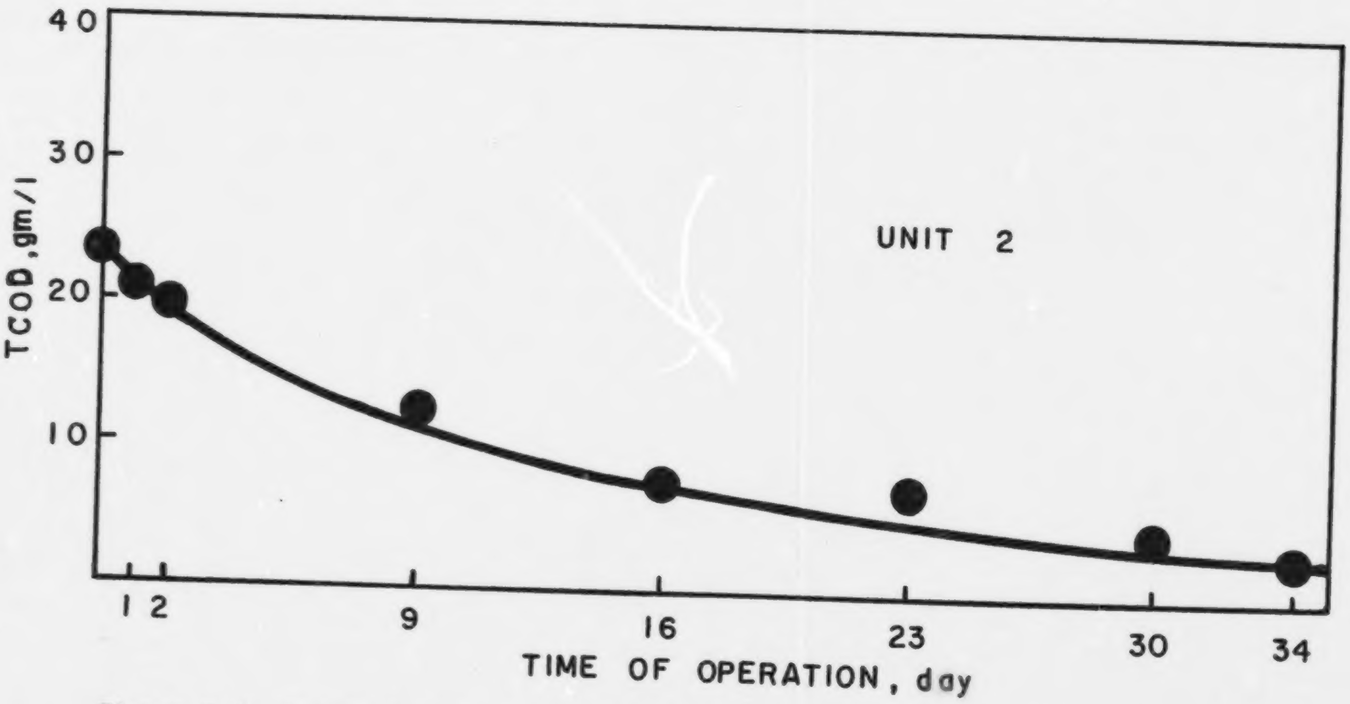

Figure C-6. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 1.

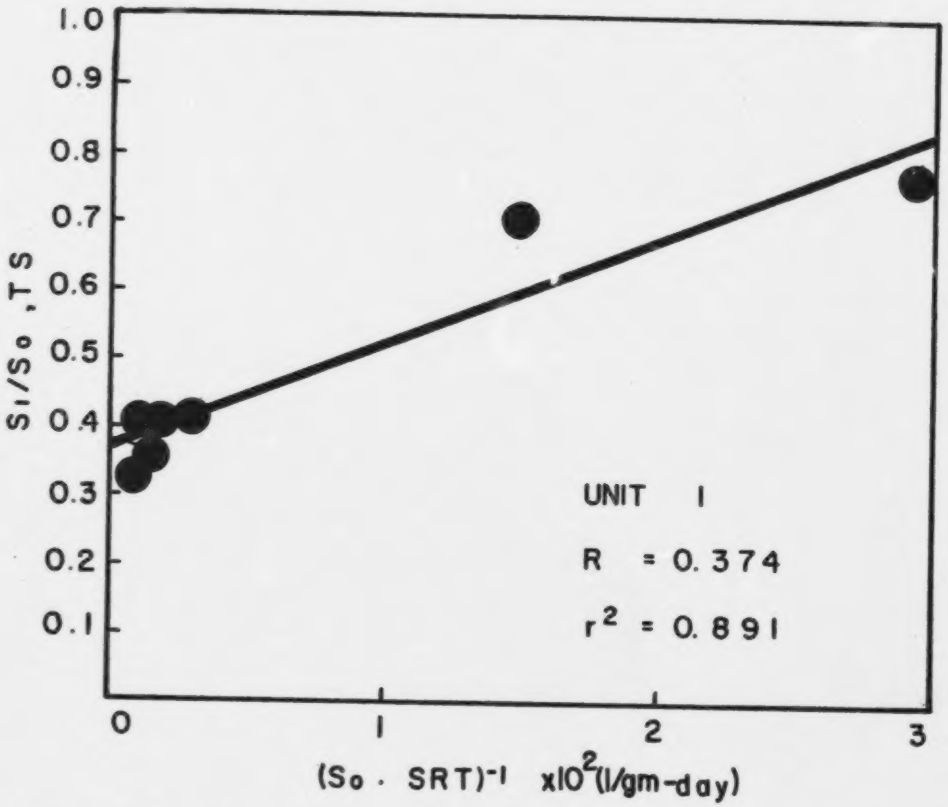

Figure C-7. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 1.

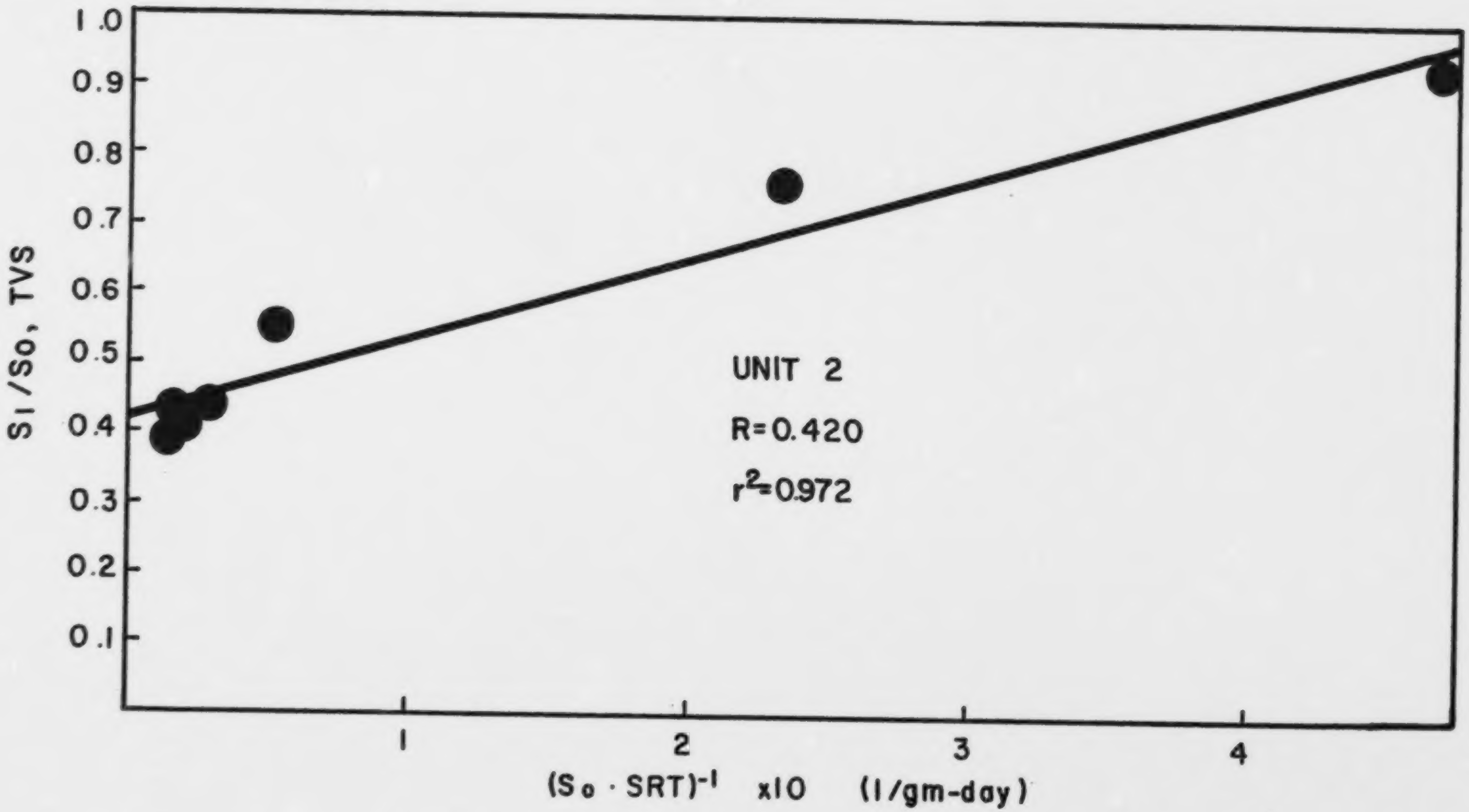

Figure C-8. Determination of the refractory fraciton (R) for total solids (TS), biodegradability study no. 1.

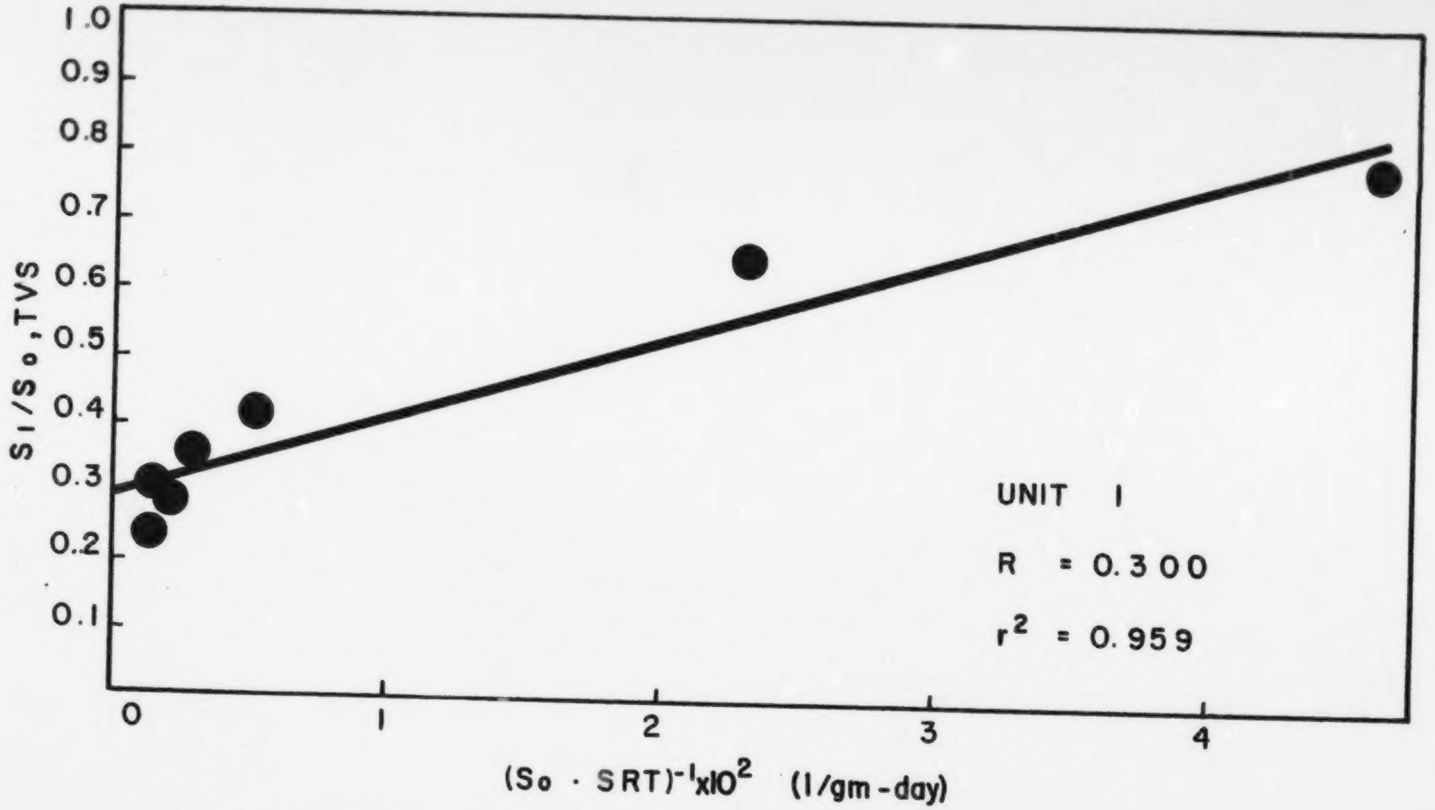

Figure C-9. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 1.

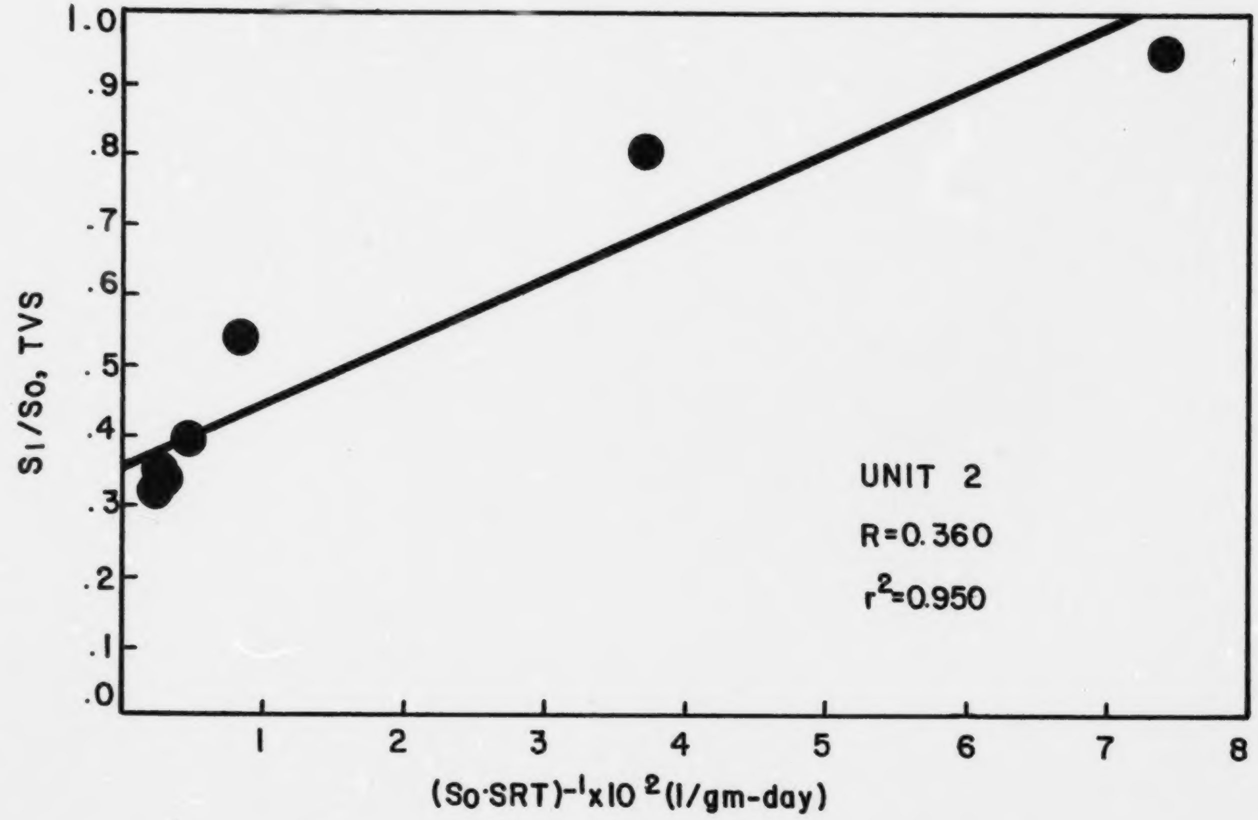

Figure C-10. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 1.

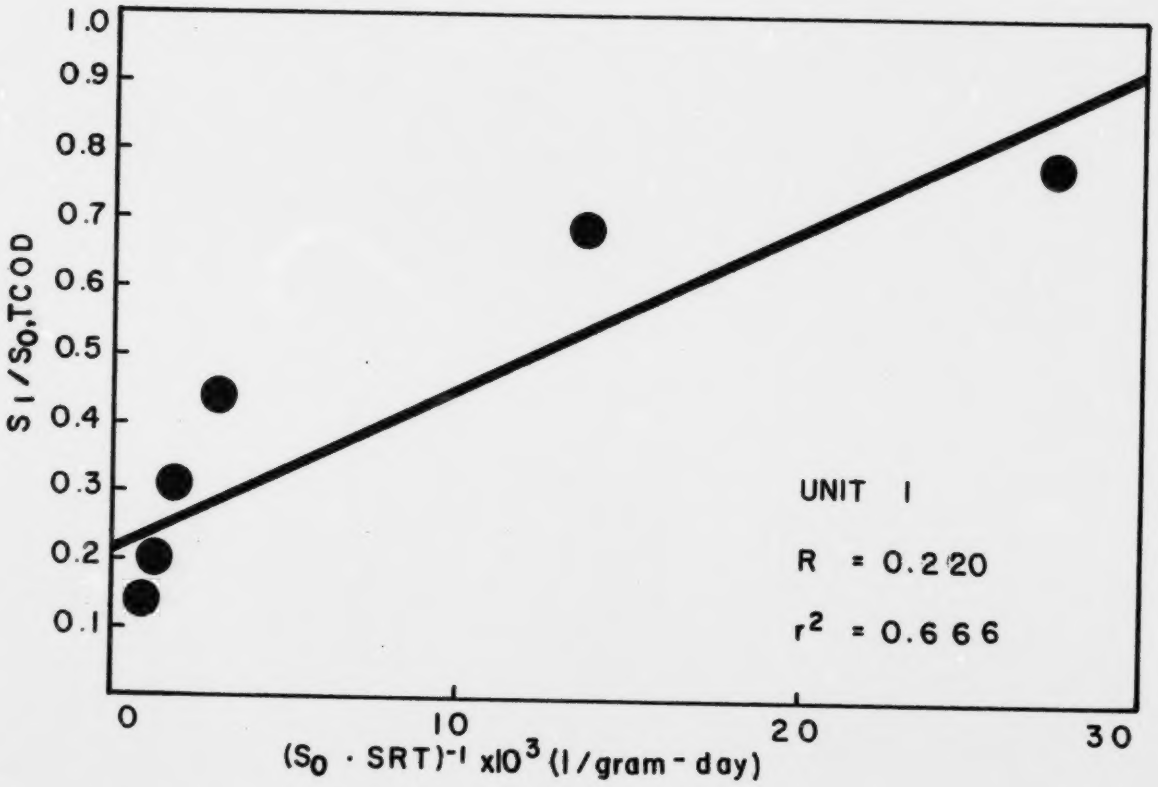

Figure C-11. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 1.

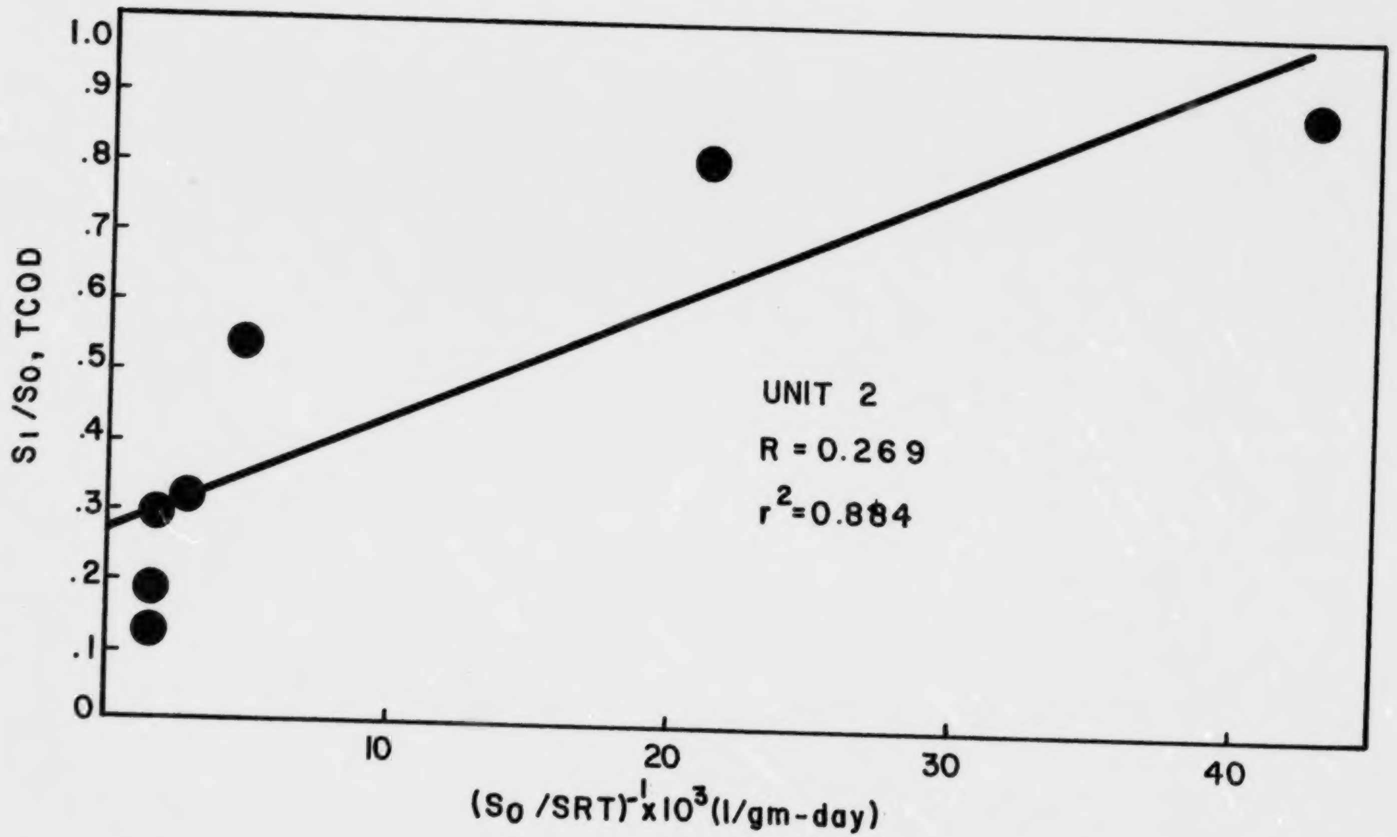

Figure C-12. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 1.

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### TABLE C-3. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 2, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 39.908 | 37.331 | 22.953 | -- | -- | -- | 61.5 |
| 1 | 34.369 | 35.213 | 21.495 | 13.9 | 5.7 | 6.4 | 61.0 |
| 2 | 34.389 | 34.185 | 20.784 | 13.9 | 8.4 | 9.4 | 60.8 |
| 9 | 16.889 | 21.434 | 11.606 | 55.6 | 42.6 | 49.4 | 54.2 |
| 16 | 8.546 | 13.360 | 6.509 | 78.0 | 64.2 | 71.6 | 48.7 |
| 23 | -- | -- | -- | -- | -- | -- | -- |
| 30 | 6.066 | 12.679 | 4.556 | 84.8 | 66.0 | 80.2 | 35.9 |

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| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 52.2 | 48-55 |
| D.O., mg/ℓ | 4.1 | 0.1-5.6 |
| pH | 7.0 | 5.9-7.3 |
| Evaporation, mℓ/day | 300 | 0-900 |
| Airflow rate, ℓ/sec | 0.08 | - |

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TABLE C-4. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 2, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 38.627 | 38.392 | 23.646 | -- | -- | -- | 61.6 |
| 1 | 37.289 | 34.953 | 22.213 | 3.5 | 9.0 | 6.1 | 63.6 |
| 2 | 35.978 | 34.430 | 20.940 | 6.9 | 10.2 | 11.5 | 60.8 |
| 9 | 16.899 | 20.981 | 12.419 | 56.3 | 45.4 | 47.5 | 59.2 |
| 16 | 9.248 | 14.957 | 7.110 | 76.1 | 61.0 | 70.0 | 47.5 |
| 23 | 6.595 | 13.167 | 5.932 | 82.9 | 65.7 | 74.9 | 45.1 |
| 30 | 3.916 | 7.689 | 3.075 | 89.9 | 80.0 | 87.0 | 40.0 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 52.1 | 49-56 |
| D.O., mg/ℓ | 4.5 | 0.1-5.6 |
| pH | 6.9 | 5.5-7.2 |
| Evaporation, mℓ/day | 480 | 0-1125 |
| Air flow rate, ℓ/sec | 0.08 | - |

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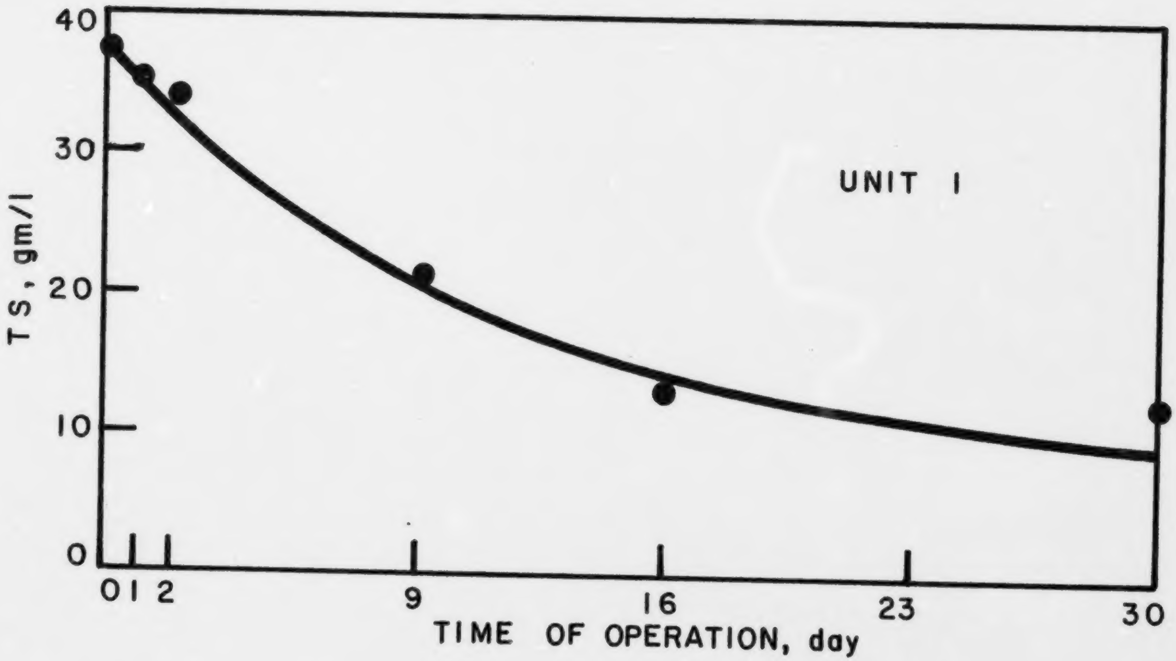

Figure C-13. Total solids (TS) removal as related to time of operation, biodegradability study no. 2.

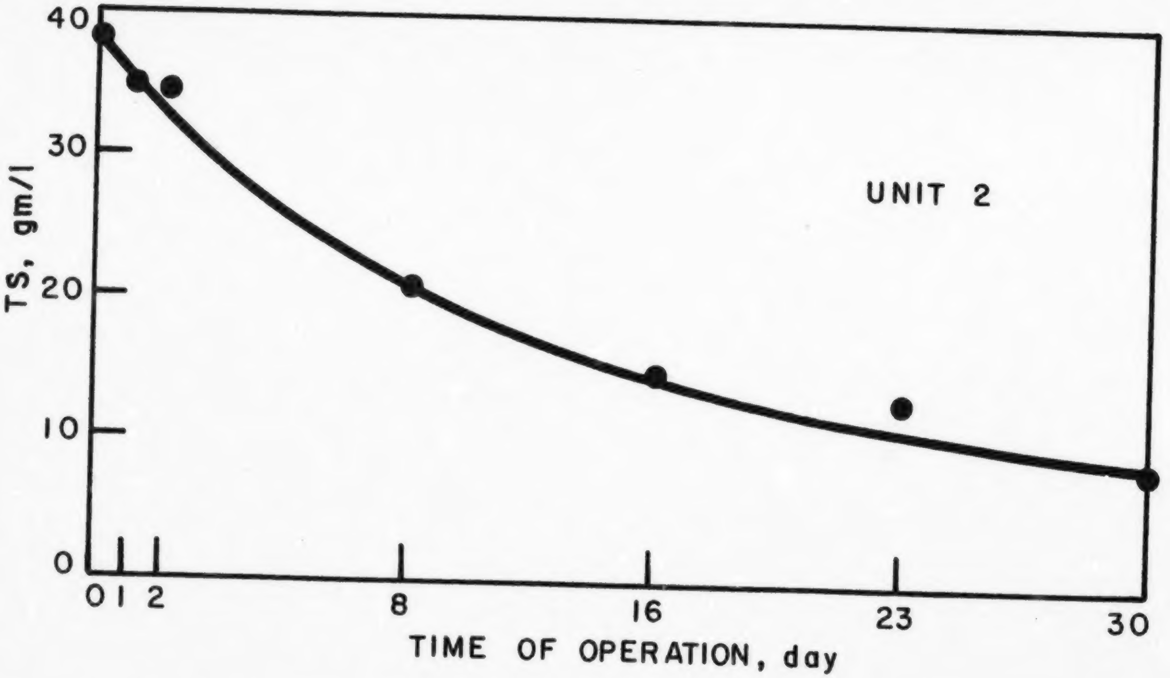

Figure C-14. Total solids (TS) removal as related to time of operation, biodegradability study no. 2.

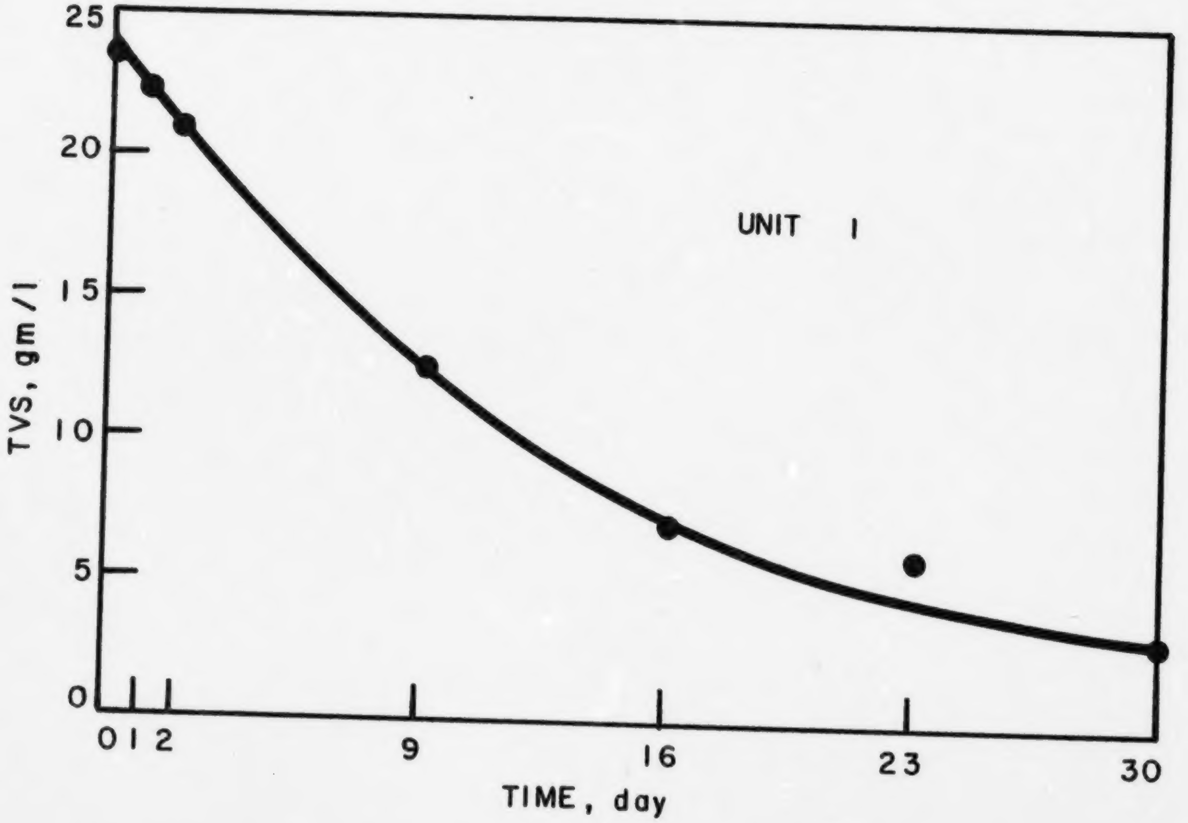

Figure C-15. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 2.

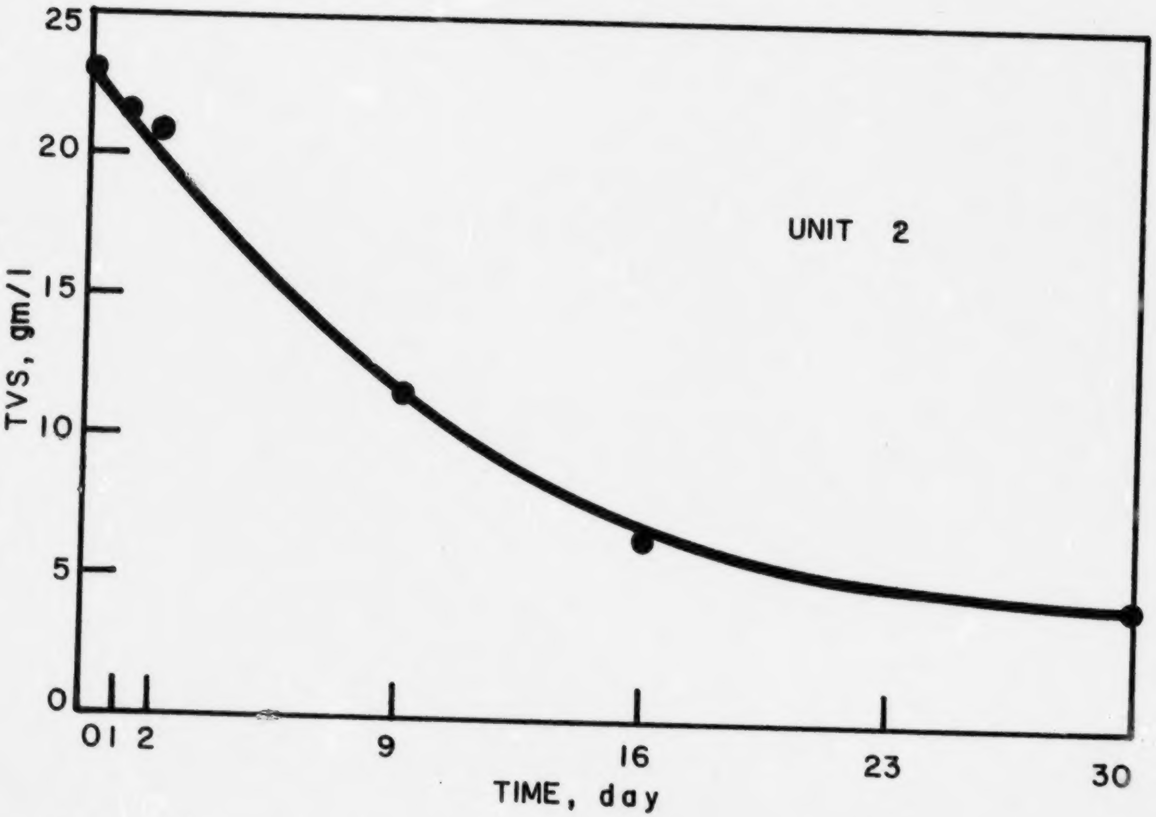

Figure C-16. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 2.

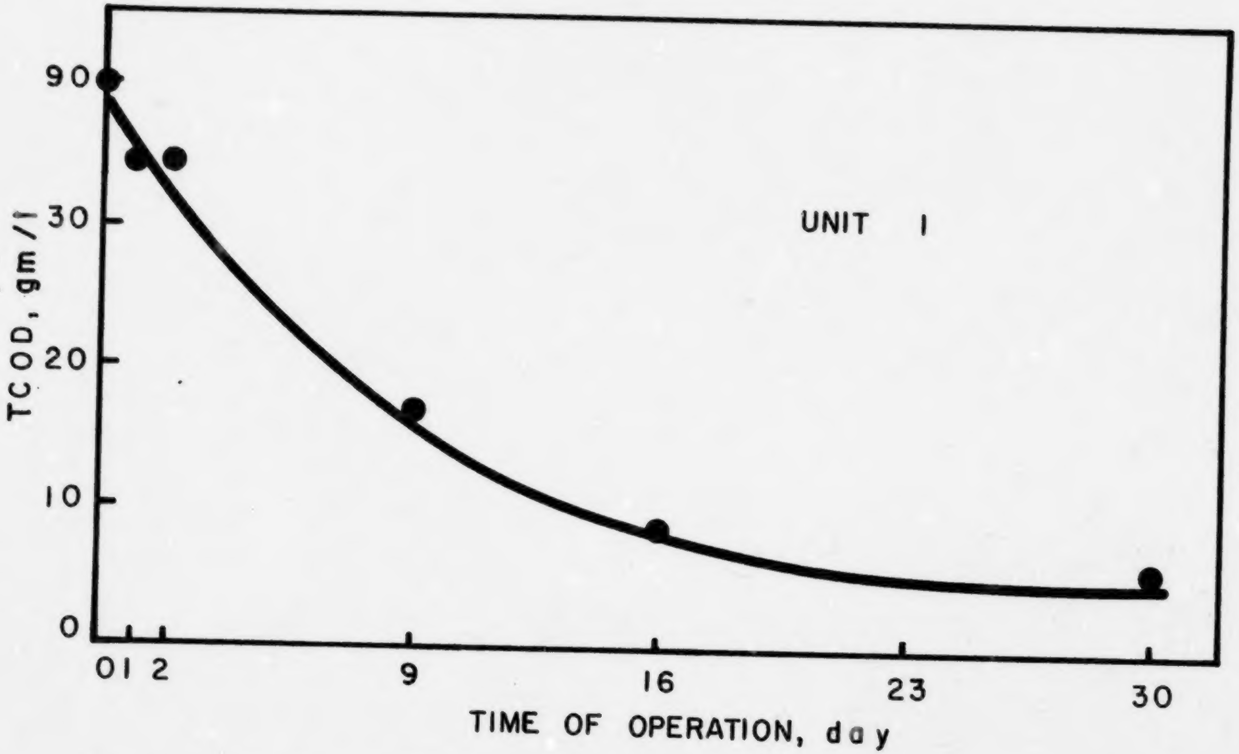

Figure C-17. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 2.

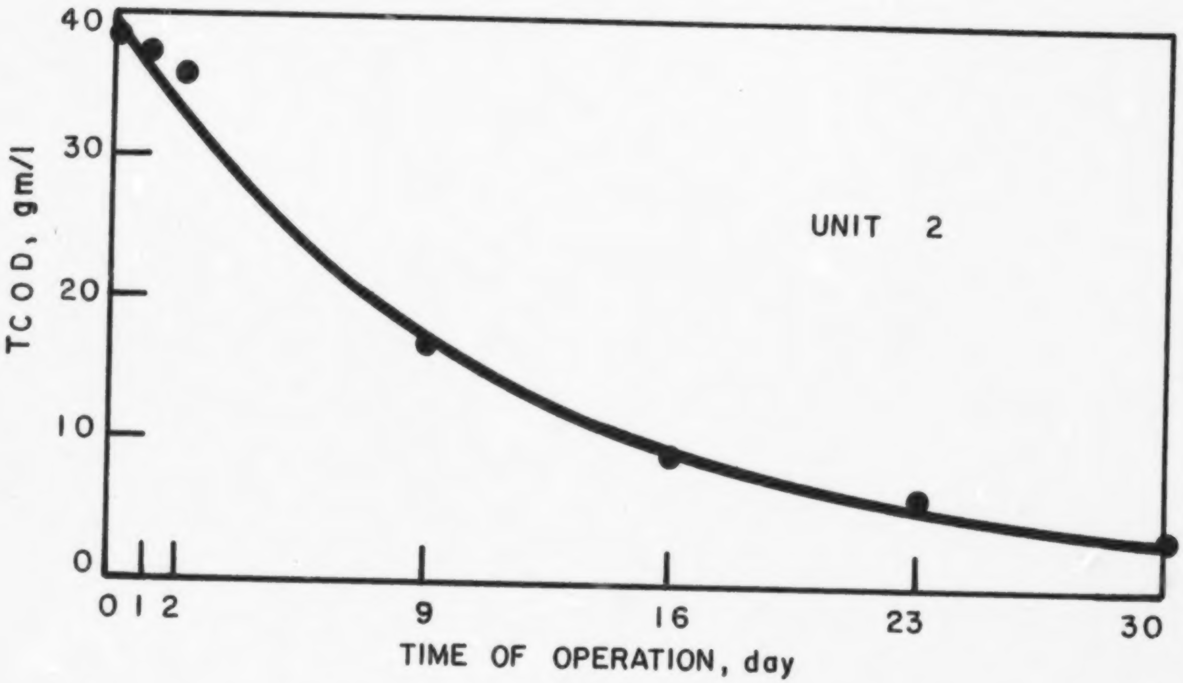

Figure C-18. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 2.

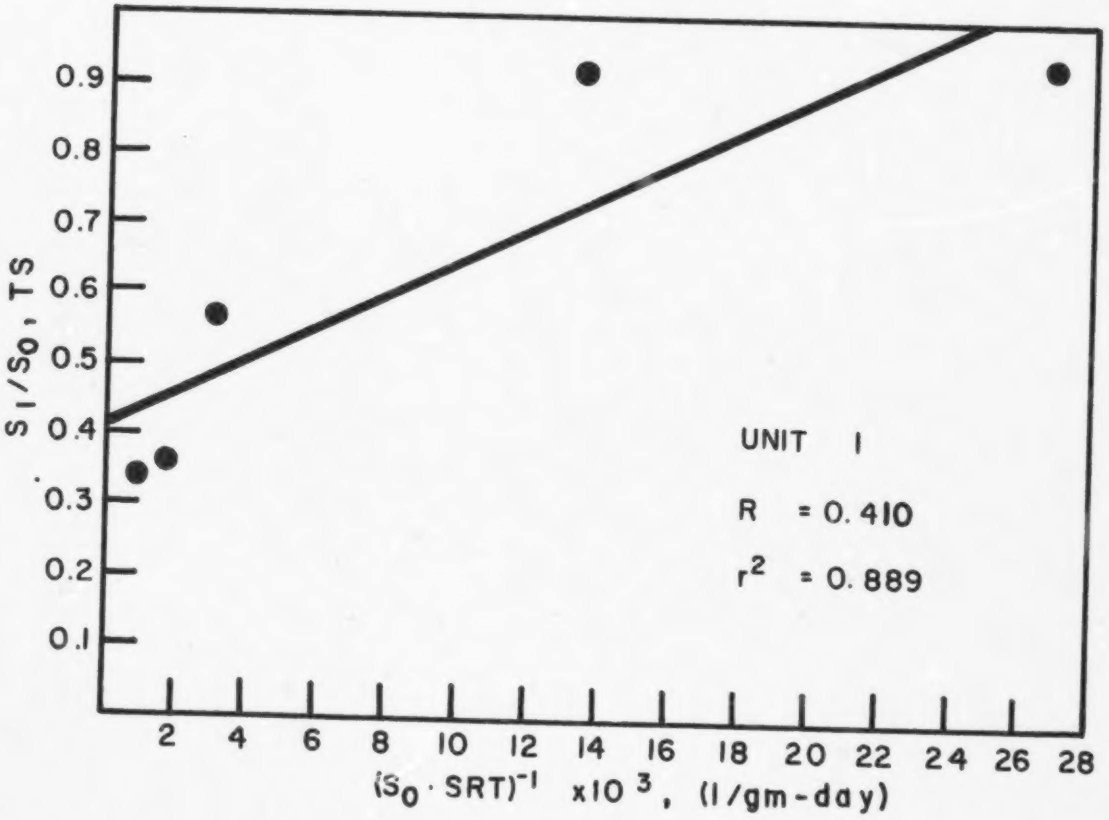

Figure C-19. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 2.

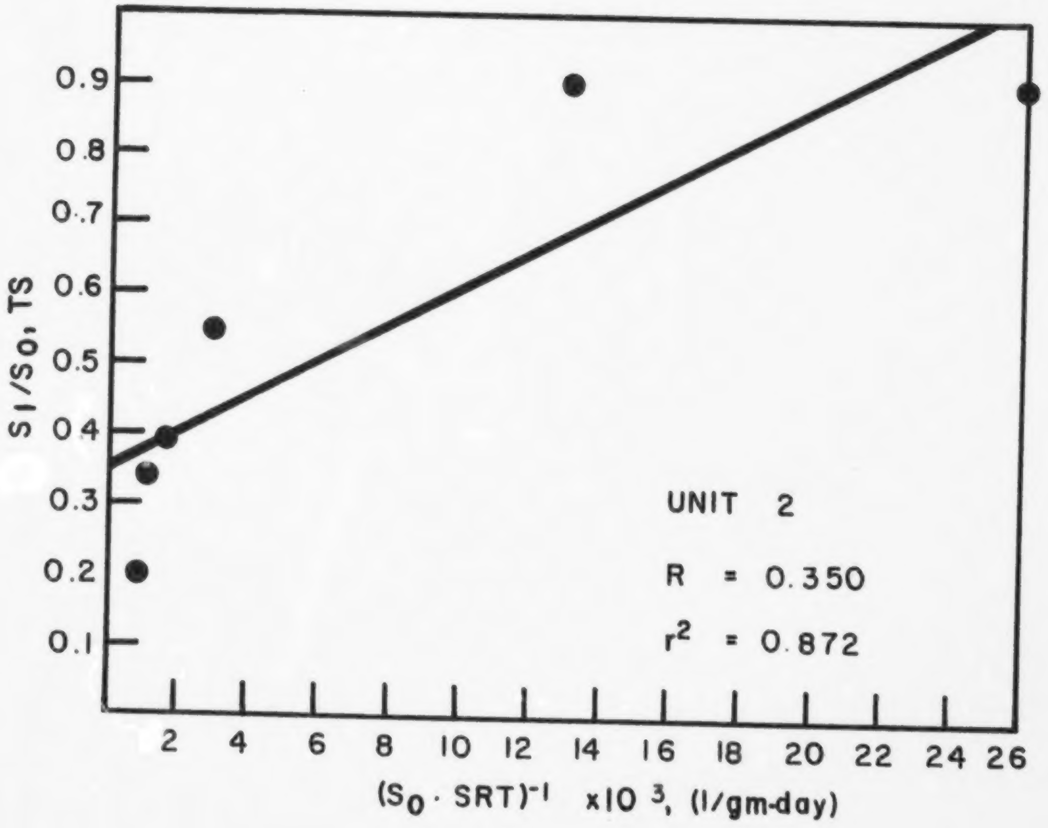

Figure C-20. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 2.

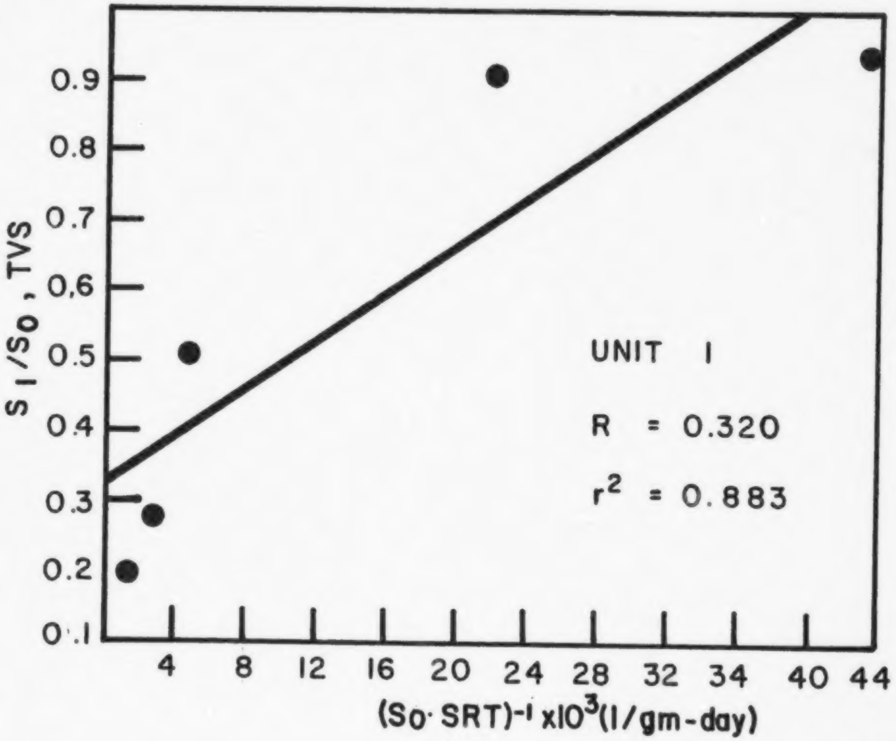

Figure C-21. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 2.

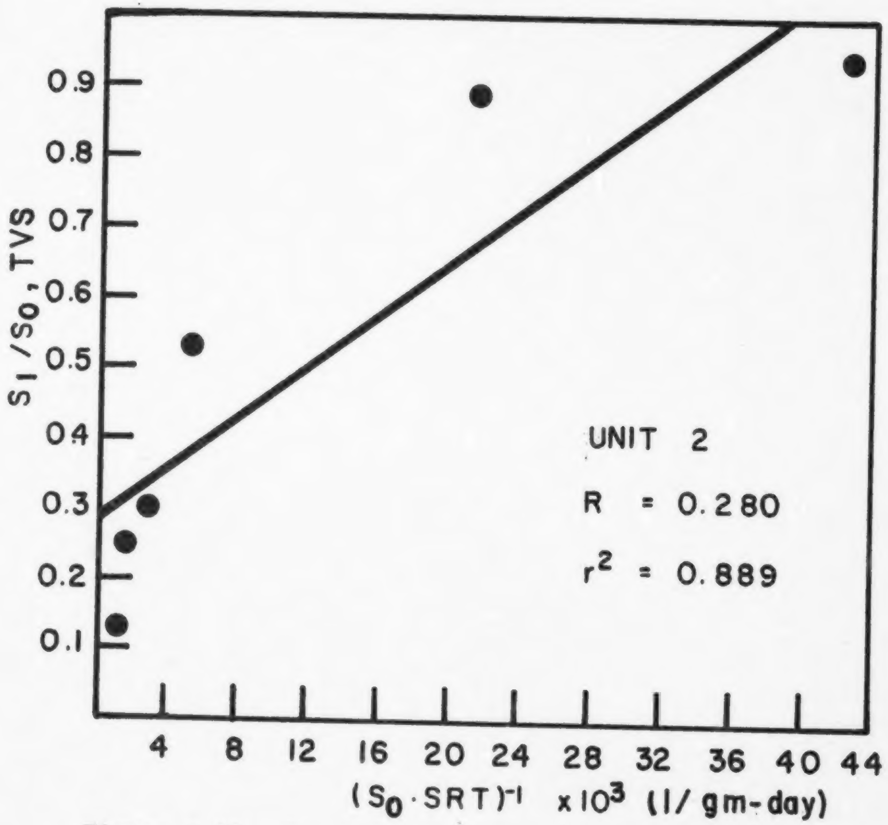

Figure C-22. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 2.

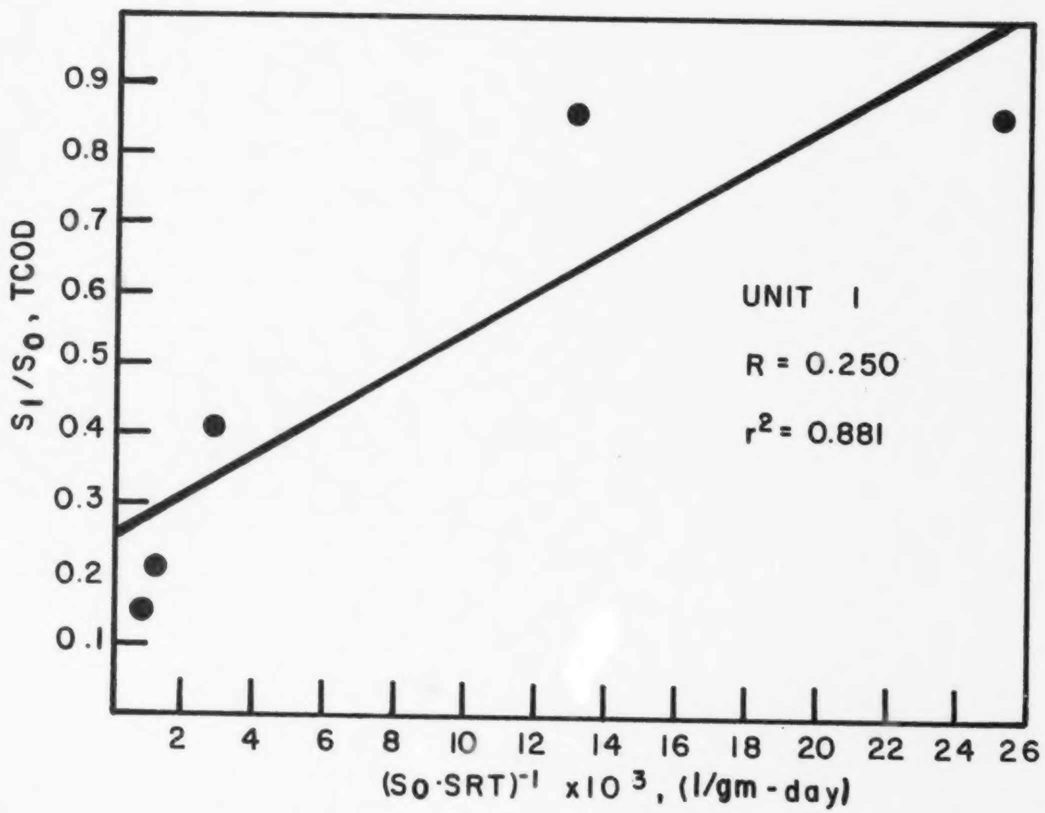

Figure C-23. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 2.

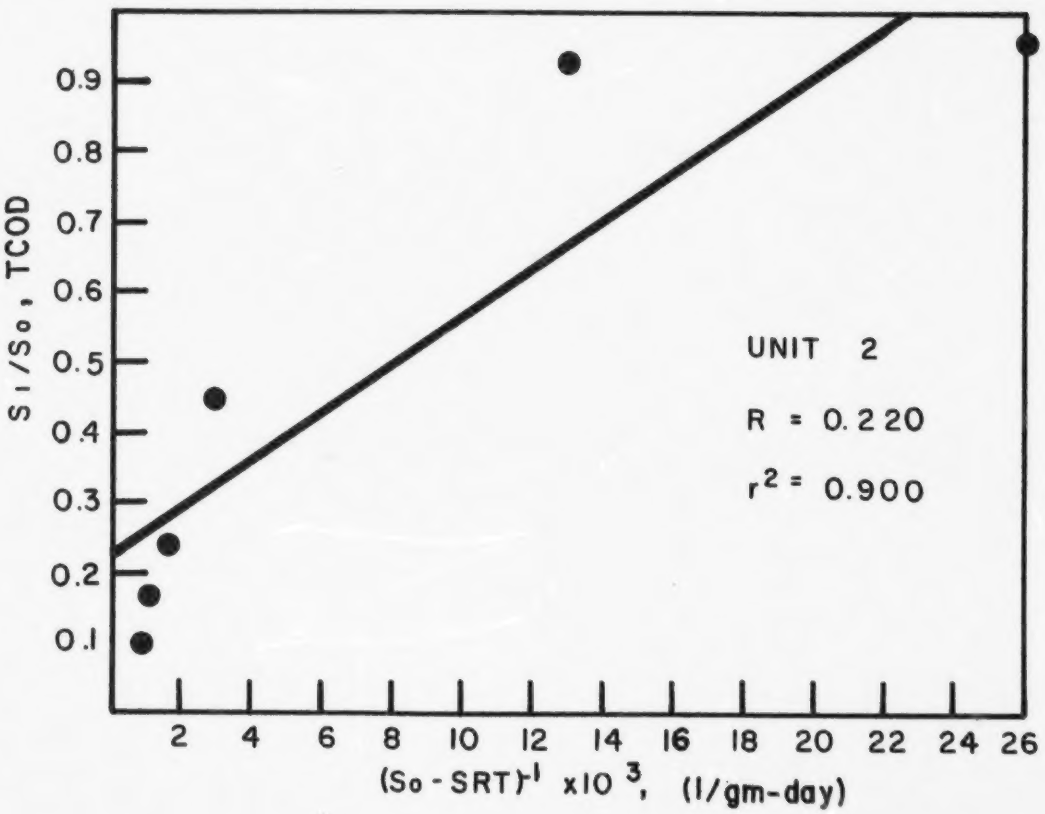

Figure C-24. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 2.

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### TABLE C-5. RESULTS OF BIODEGRADABILITY STUDY NO. 3, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 42.113 | 43.878 | 25.694 | -- | -- | -- | 58.6 |
| 1 | 39.935 | 36.645 | 23.099 | 5.2 | 12.2 | 10.1 | 63.1 |
| 2 | 33.079 | 35.052 | 21.145 | 21.5 | 20.1 | 17.7 | 60.3 |
| 6 | 20.818 | 26.589 | 15.550 | 50.6 | 39.4 | 39.5 | 58.5 |
| 10 | 14.689 | 16.169 | 9.977 | 65.1 | 63.2 | 61.2 | 61.7 |
| 17 | 12.031 | 19.292 | 8.190 | 71.4 | 56.0 | 68.1 | 42.5 |
| 26 | 10.593 | 19.470 | 7.940 | 74.8 | 55.6 | 69.1 | 40.8 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 54 | 38-55 |
| D.O., mg/ℓ | 2.5 | 0.3-4.4 |
| pH | 7.0 | 5.8-7.3 |
| Evaporation, mℓ/day | 200 | 0-925 |
| Airflow rate, ℓ/sec | 0.08 | - |

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## TABLE C-6. RESULTS OF BIODEGRADABILITY STUDY NO. 3, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 42.130 | 43.834 | 26.005 | -- | -- | -- | 59.3 |
| 1 | 39.337 | 36.340 | 21.845 | 6.6 | 17.1 | 16.0 | 60.1 |
| 2 | 33.656 | 33.254 | 20.451 | 20.1 | 24.1 | 21.4 | 61.5 |
| 6 | 29.138 | 33.067 | 19.096 | 34.9 | 24.6 | 26.6 | 57.6 |
| 10 | 14.693 | 18.611 | 10.418 | 65.1 | 57.5 | 59.9 | 56.0 |
| 17 | 10.559 | 15.842 | 7.668 | 74.9 | 63.9 | 70.5 | 48.4 |
| 26 | 9.534 | 13.819 | 5.978 | 77.4 | 68.5 | 77.0 | 43.3 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 53 | 45-61 |
| D.O., mg/ℓ | 1.5 | 0.1-3.4 |
| pH | 7.0 | 5.6-7.4 |
| Evaporation, mℓ/day | 500 | 50-1000 |
| Air flow rate, ℓ/sec | 0.08 | - |

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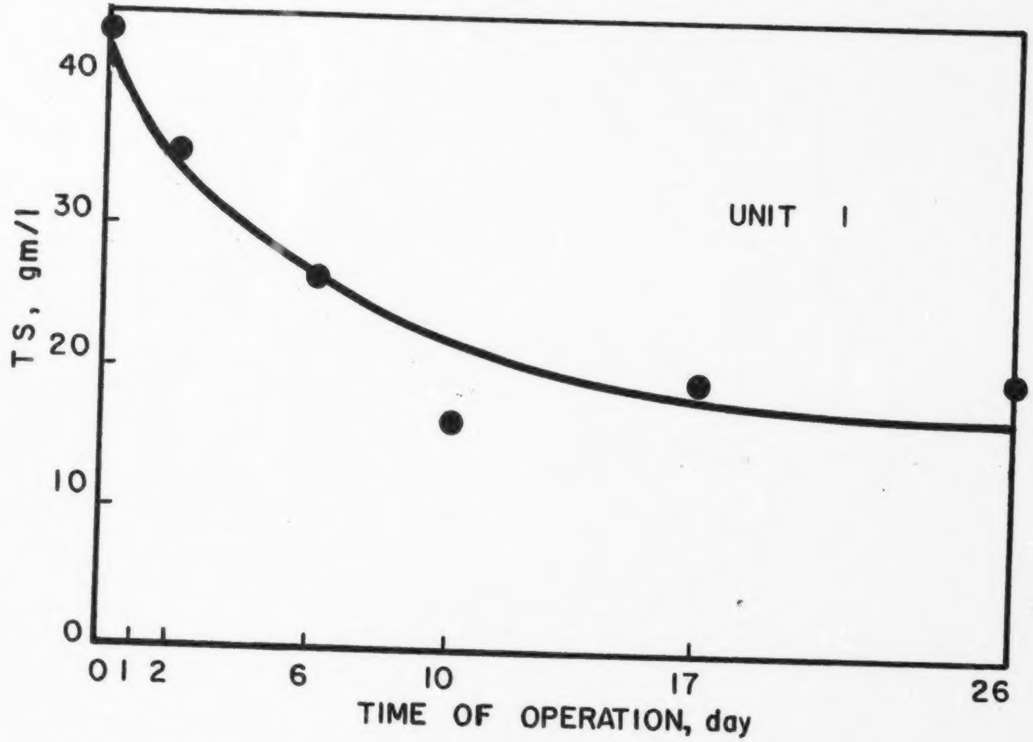

Figure C-25. Total solids (TS) removal as related to time of operation, biodegradability study no. 3.

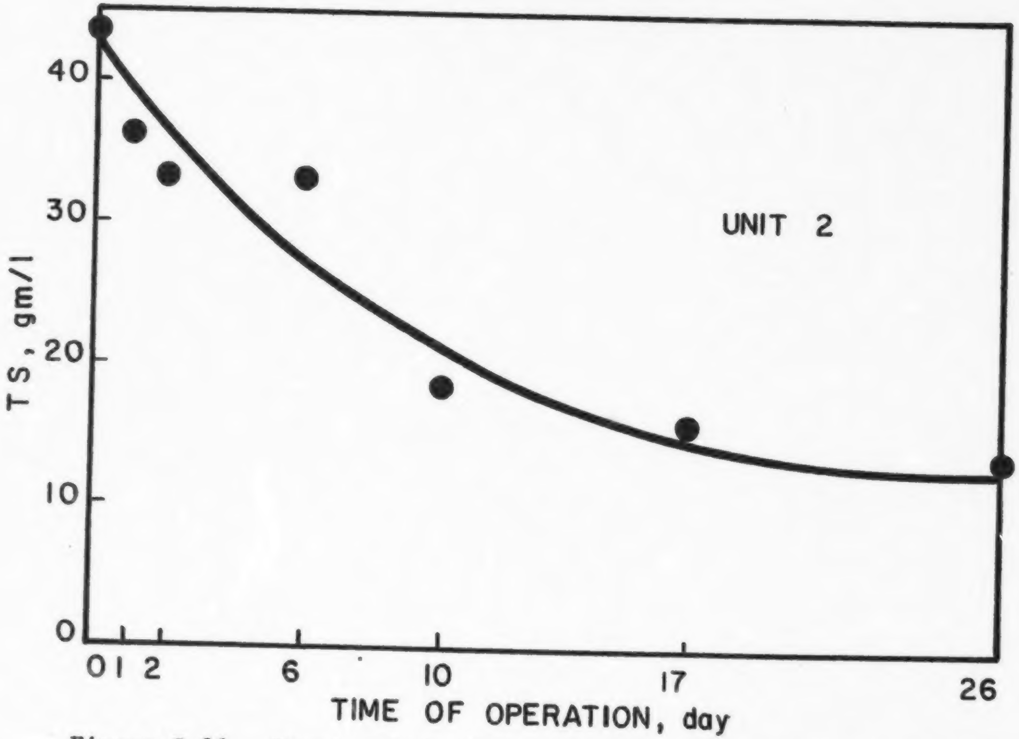

Figure C-26. Total solids (TS) removal as related to time of operation, biodegradability study no. 3.

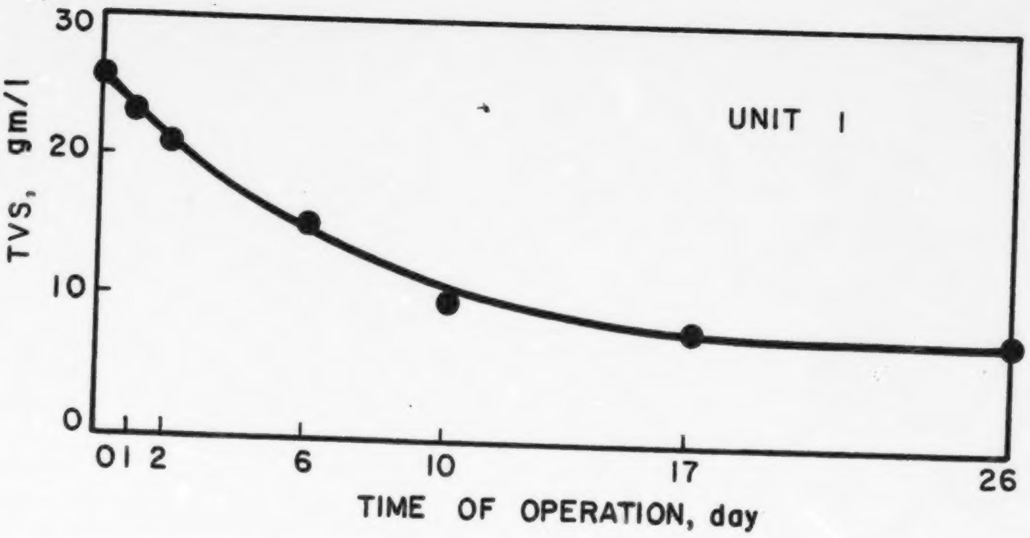

Figure C-27. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 3.

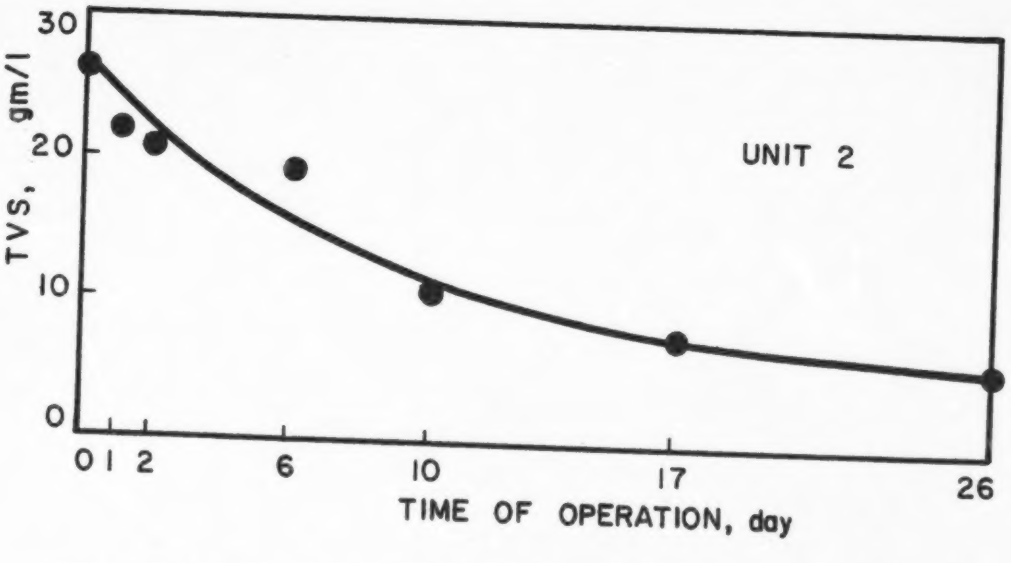

Figure C-28. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 3.

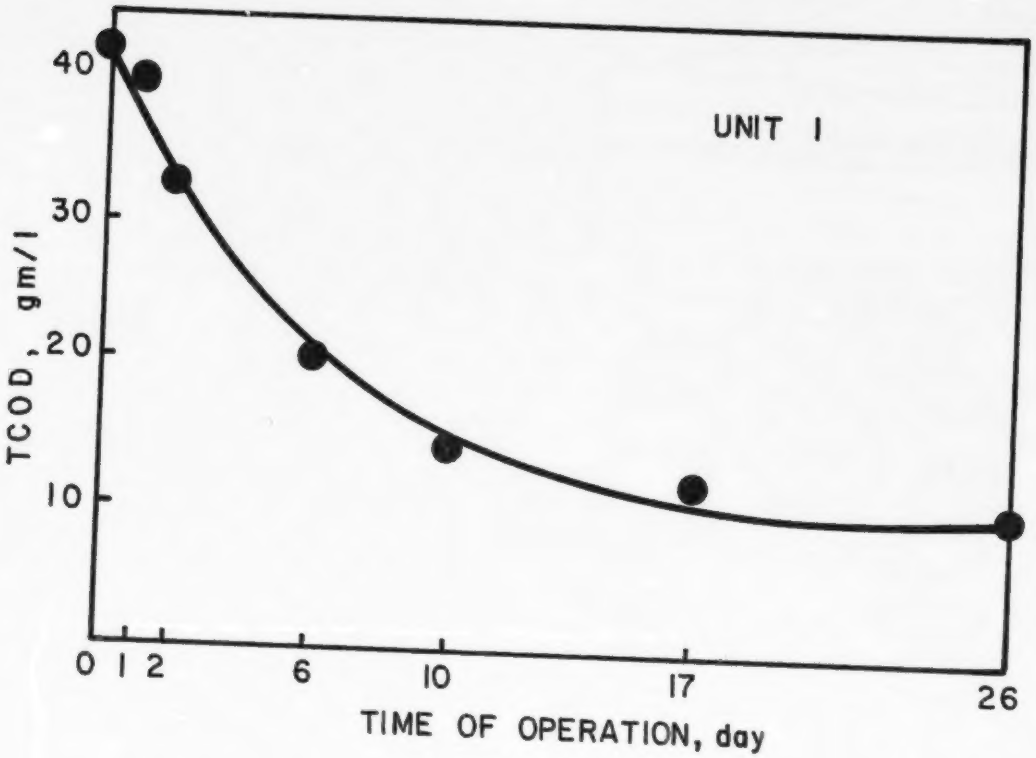

Figure C-29. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 3.

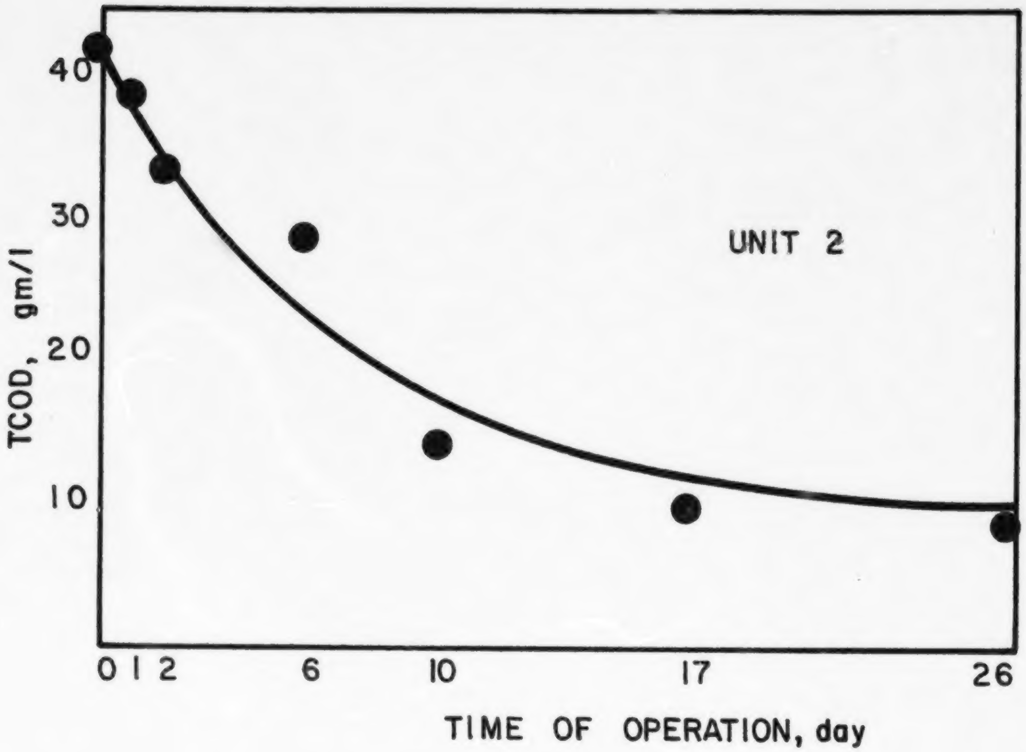

Figure C-30. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 3.

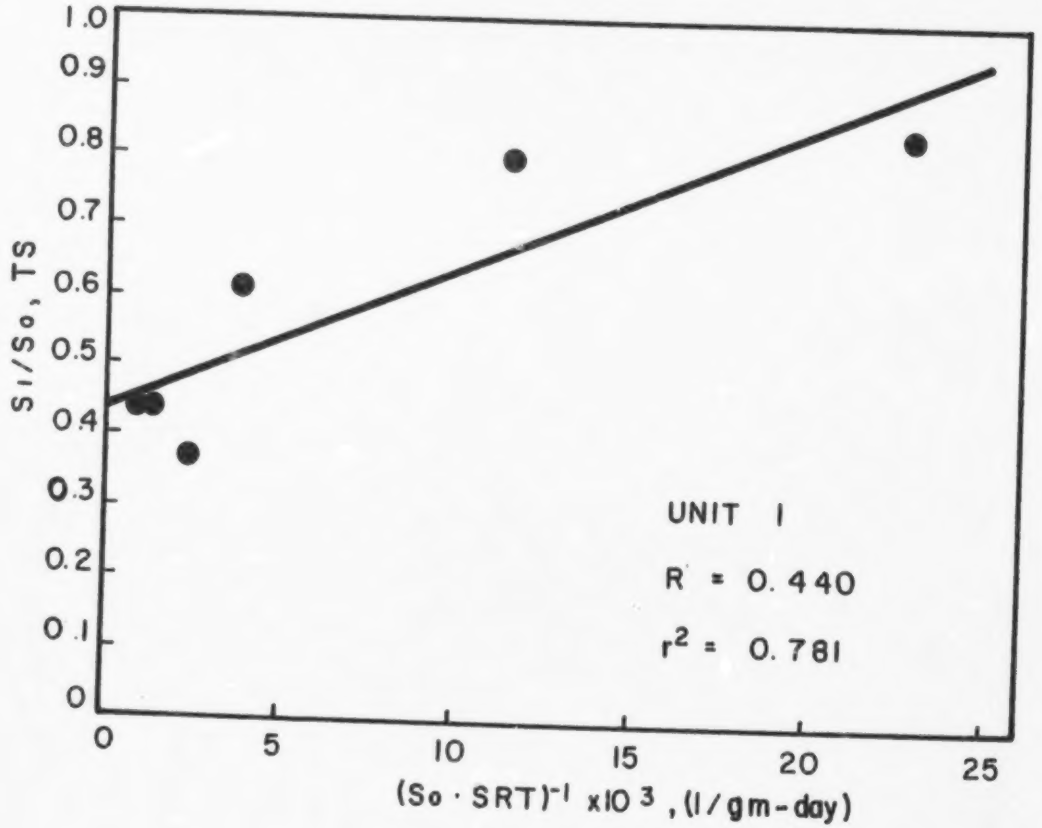

Figure C-31. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 3.

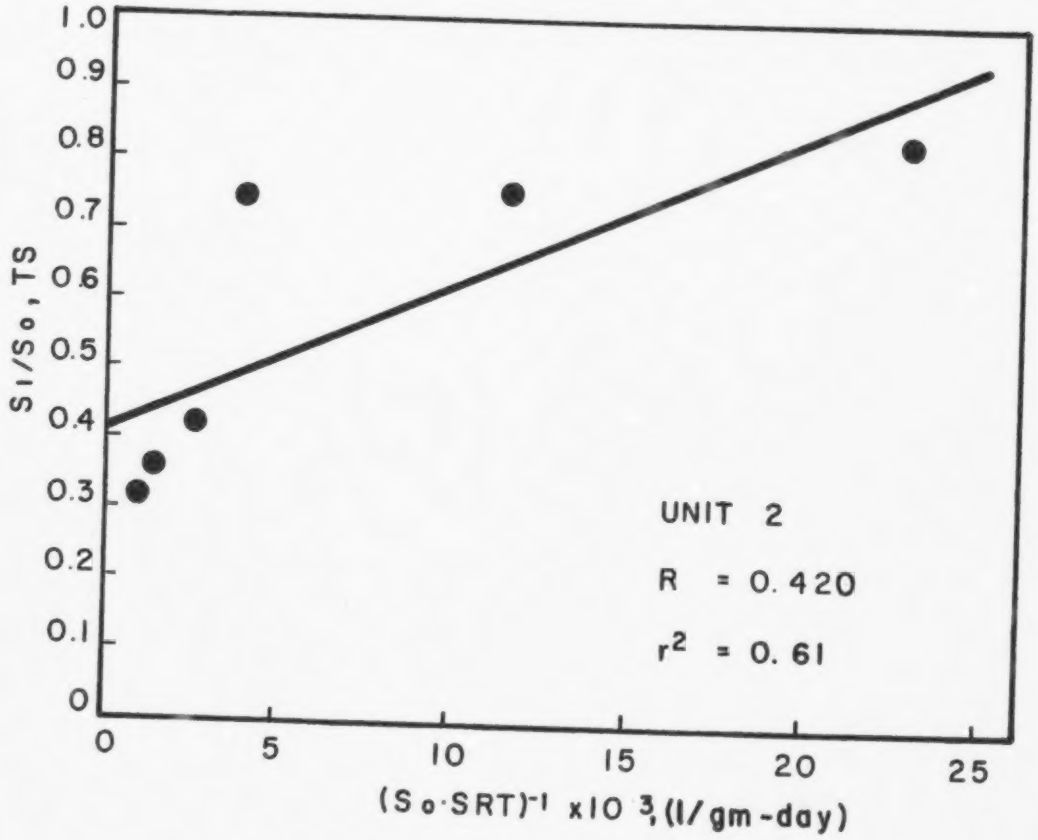

Figure C-32. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 3.

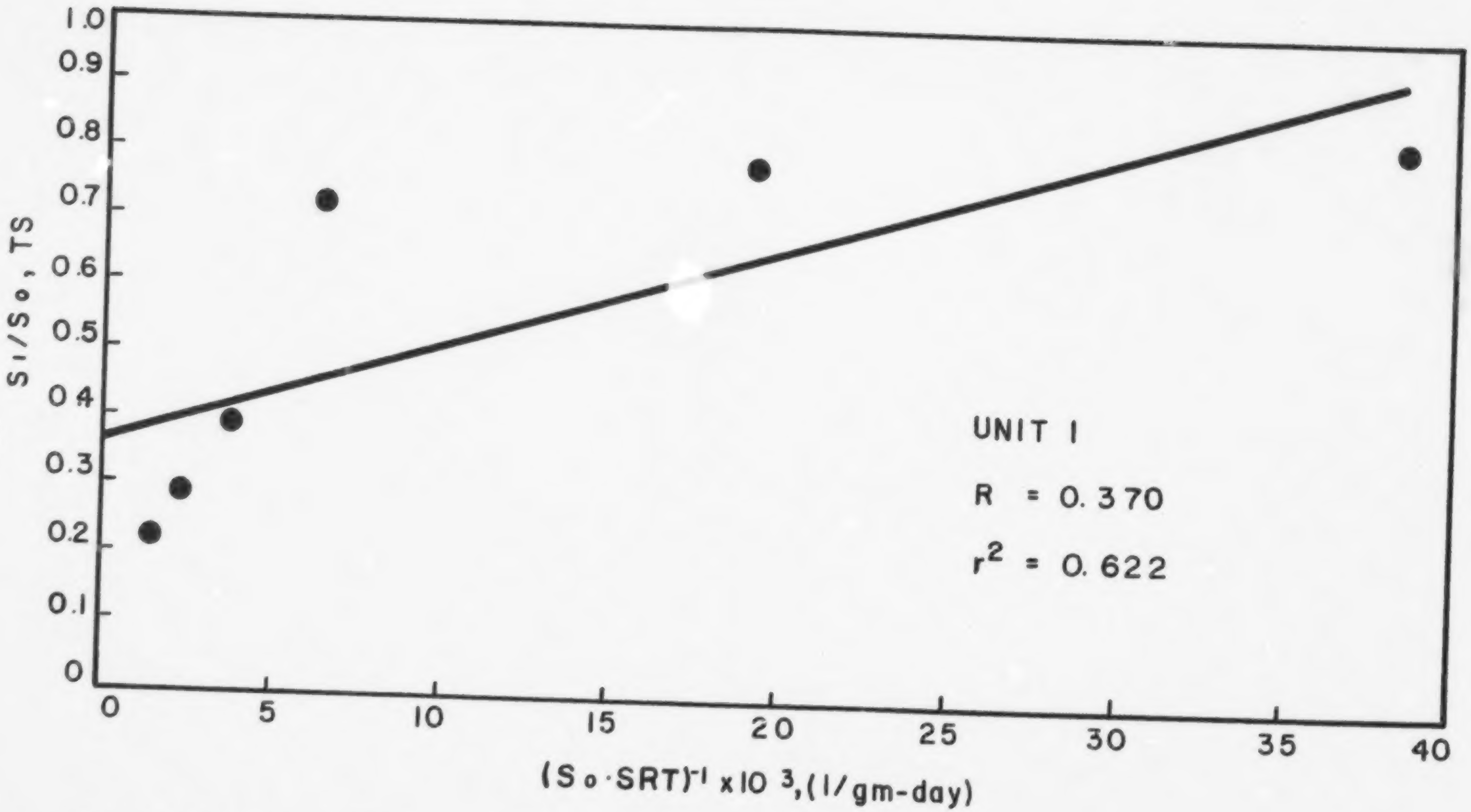

Figure C-33. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 3.

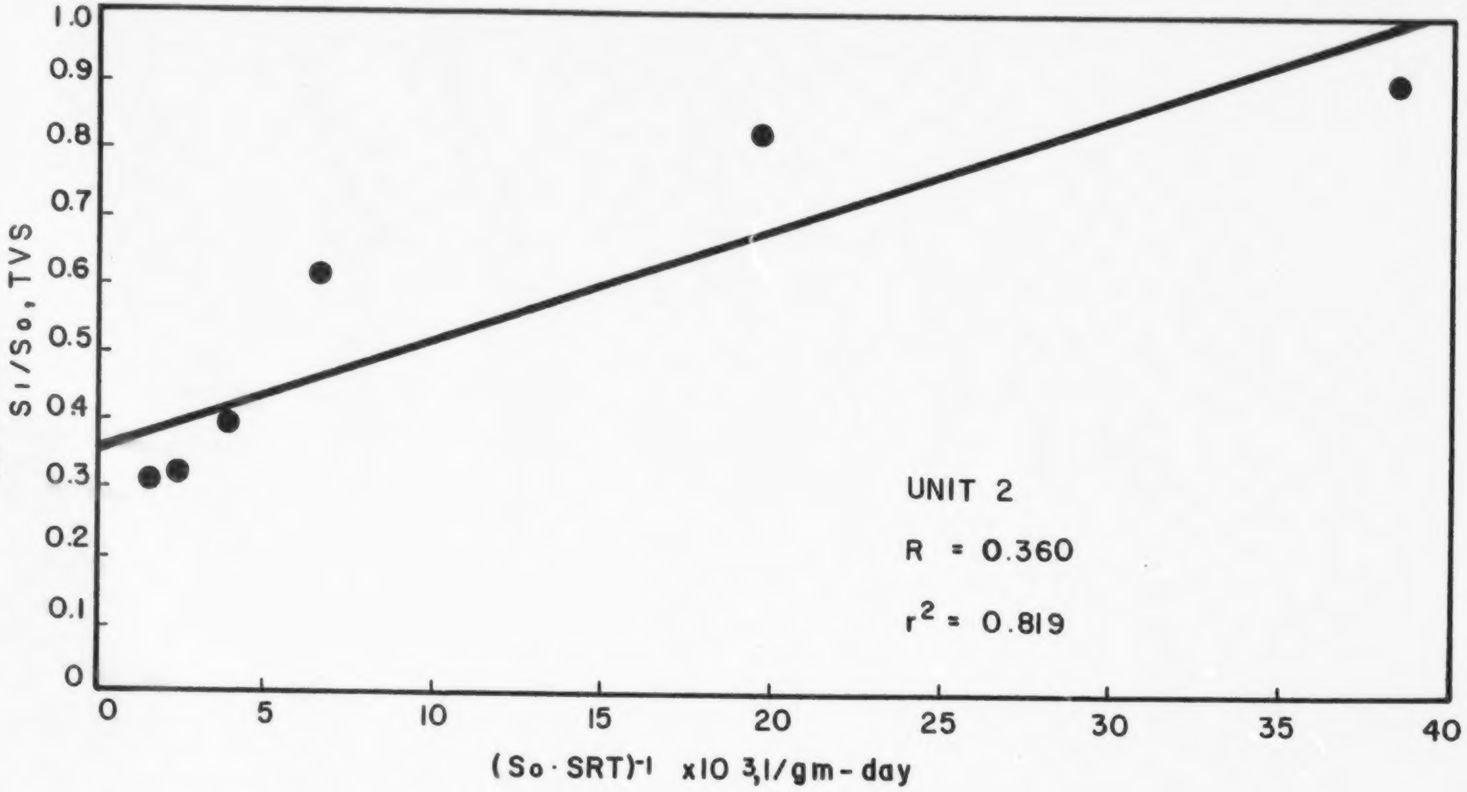

Figure C-34. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 3.

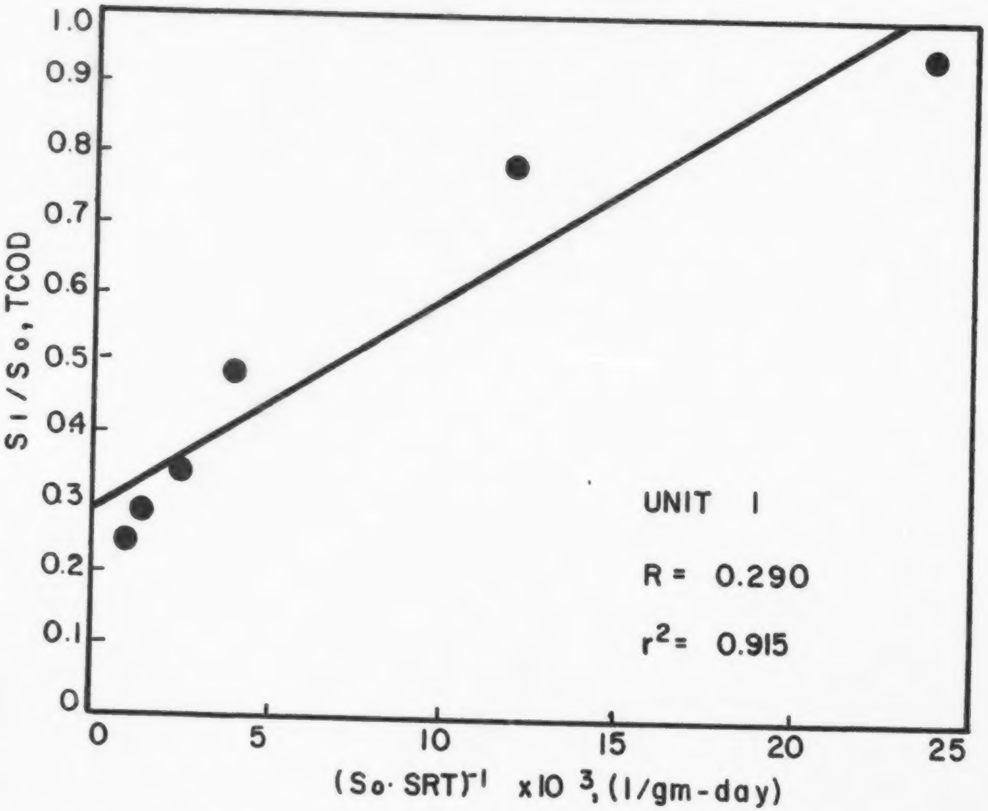

Figure C-35. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 3.

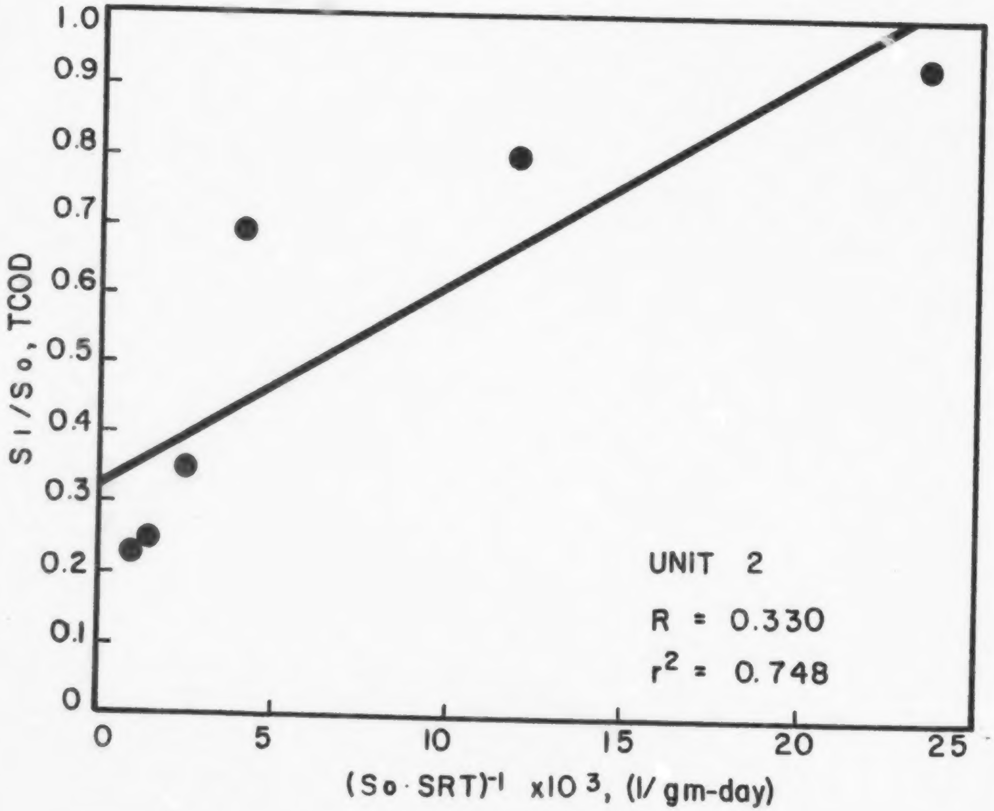

Figure C-36. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 3.

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TABLE C-7. RESULTS OF BIODEGRADABILITY STUDY NO. 4, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | TKN mg/ℓ | $NH_3$-N mg/ℓ | pH | OUR mg/gmVS/hr |
|---|---|---|---|---|---|---|---|
| 0 | 44.802 | 41.438 | 28.583 | 2247.0 | 138.6 | 5.75 | -- |
| 1 | 38.095 | 36.058 | 24.037 | -- | -- | 7.30 | -- |
| *2 | 26.020 | 22.884 | 14.882 | -- | -- | 7.10 | -- |
| *3 | 18.357 | 20.116 | 12.877 | -- | -- | 7.10 | -- |
| *9 | 16.469 | 22.485 | 12.949 | -- | -- | 6.70 | 31.8 |
| 16 | 18.063 | 25.803 | 12.960 | 742.0 | 35.0 | 6.70 | 25.0 |
| 23 | 7.321 | 15.598 | 8.023 | -- | -- | 7.40 | -- |
| 30 | 11.313 | 19.314 | 7.731 | 504.0 | 8.40 | 7.50 | 34.2 |
| 37 | 9.792 | 24.478 | 10.135 | 336.0 | 5.60 | 7.60 | -- |

*Reactor cracked, allowing water from bath to enter and thereby dilute mixed liquor.

| Operating Conditions for Above Data | |
|---|---|
| Reactor temperature | 50°C ± 3°C |
| Reactor D.O. | 4.1 mg/ℓ ± 1.5 mg/ℓ |
| Evaporation | 600 mℓ/day |
| Airflow rate | 0.07 ℓ/sec |

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TABLE C-8. RESULTS OF BIODEGRADABILITY STUDY NO. 4, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | TKN mg/ℓ | $NH_3$-N mg/ℓ | pH | OUR mg/gmVS/hr |
|---|---|---|---|---|---|---|---|
| 0 | 44.802 | 41.438 | 28.583 | 2247.0 | 138.6 | 5.75 | -- |
| 1 | 38.693 | 37.562 | 24.671 | -- | -- | 7.25 | -- |
| 2 | 31.875 | 31.883 | 20.290 | -- | -- | 6.70 | -- |
| 3 | 29.619 | 29.815 | 18.716 | -- | -- | 6.70 | -- |
| 9 | 14.763 | 21.518 | 11.396 | -- | -- | 6.40 | 18.1 |
| 16 | 14.887 | 20.066 | 9.523 | 854.0 | 33.6 | 6.60 | 18.9 |
| 23 | 3.983 | 5.498 | 3.237 | -- | -- | 7.10 | -- |
| 30 | 4.0671 | 7.8247 | 3.790 | 287.0 | 10.5 | 7.20 | -- |
| 37 | 18.230 | 21.446 | 8.430 | 336.0 | 5.6 | 7.40 | -- |

| Operating Conditions for Above Data | |
|---|---|
| Reactor temperature | 52°C ± 3°C |
| Reactor D.O. | 3.4 mg/ℓ ± 1.6 mg/ℓ |
| Evaporation | 600 mℓ/day |
| Airflow rate | 0.07 ℓ/sec |

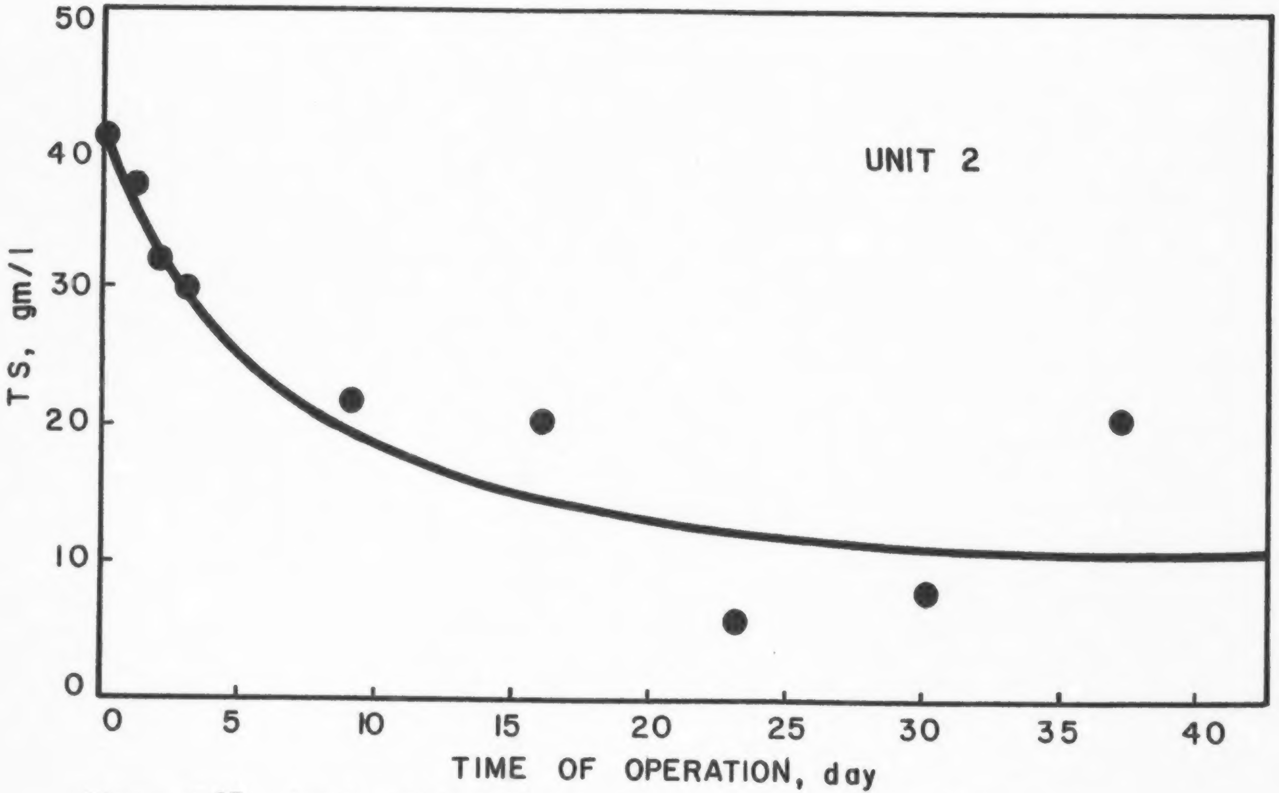

Figure C-37. Total solids (TS) removal as related to time of operation. biodegradability study no. 4.

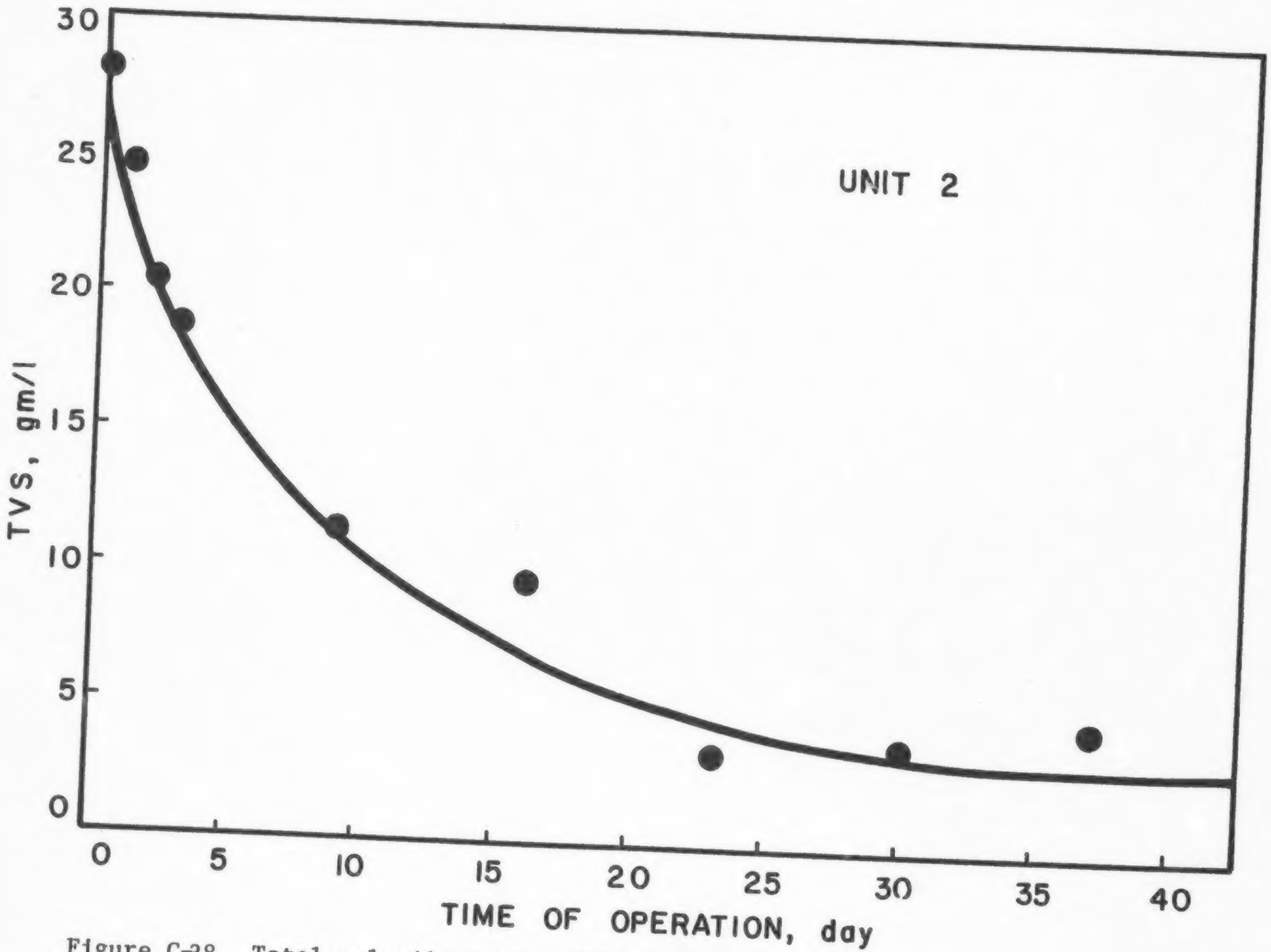

Figure C-38. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 4.

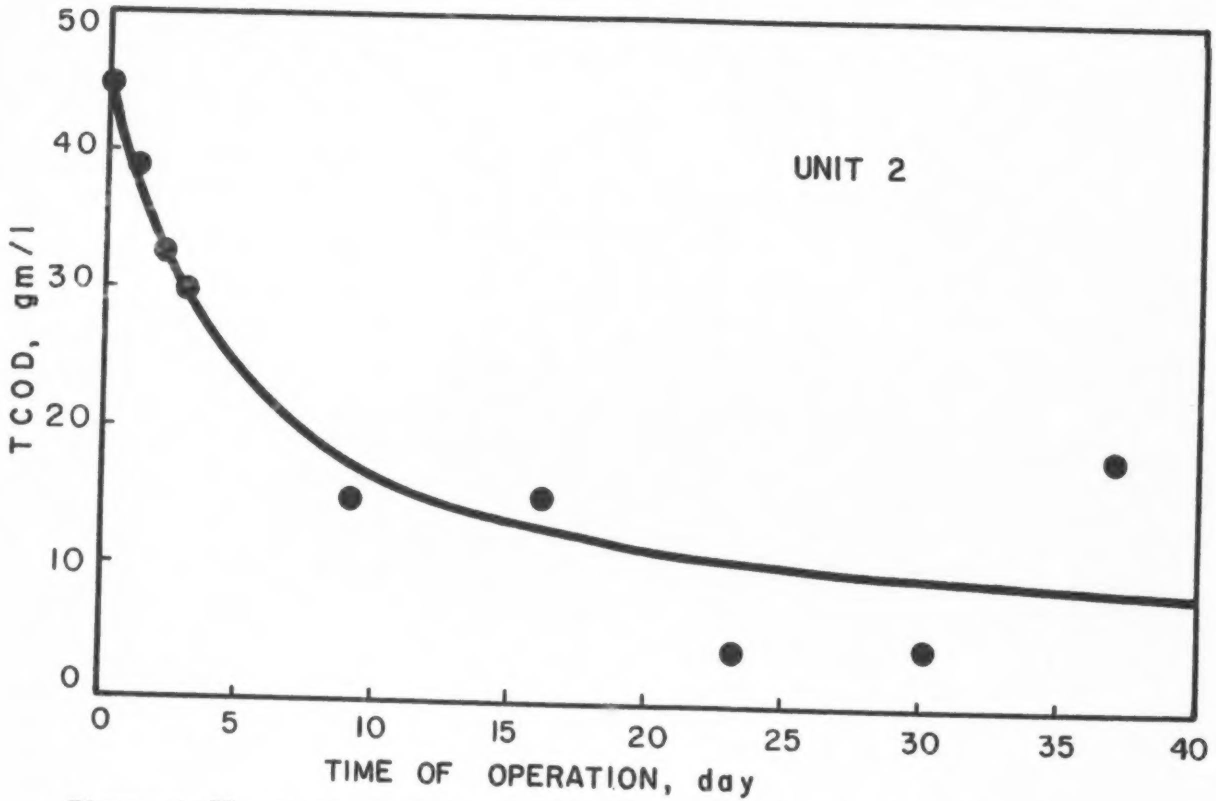

Figure C -39. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 4.

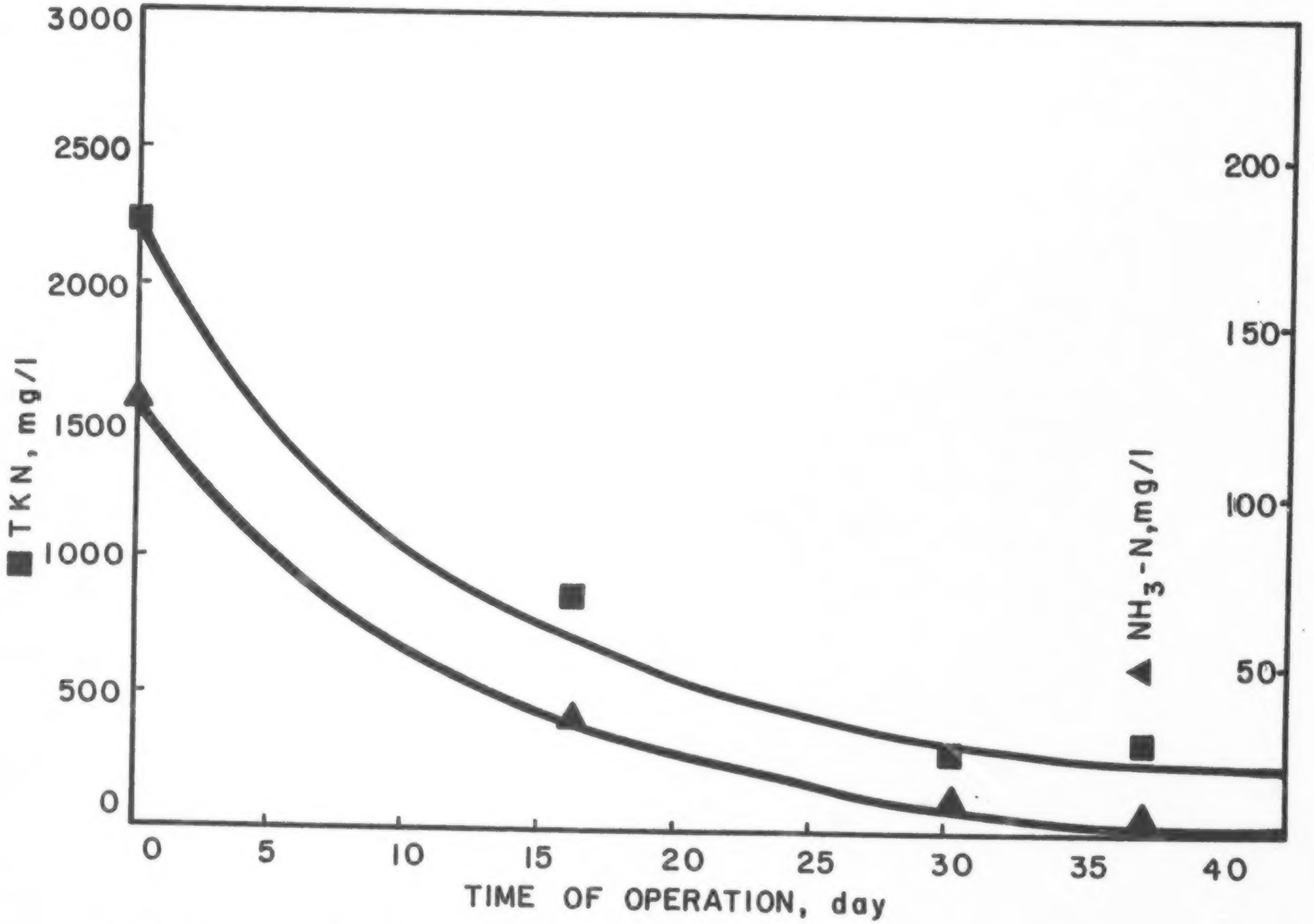

Figure C-40. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 4.

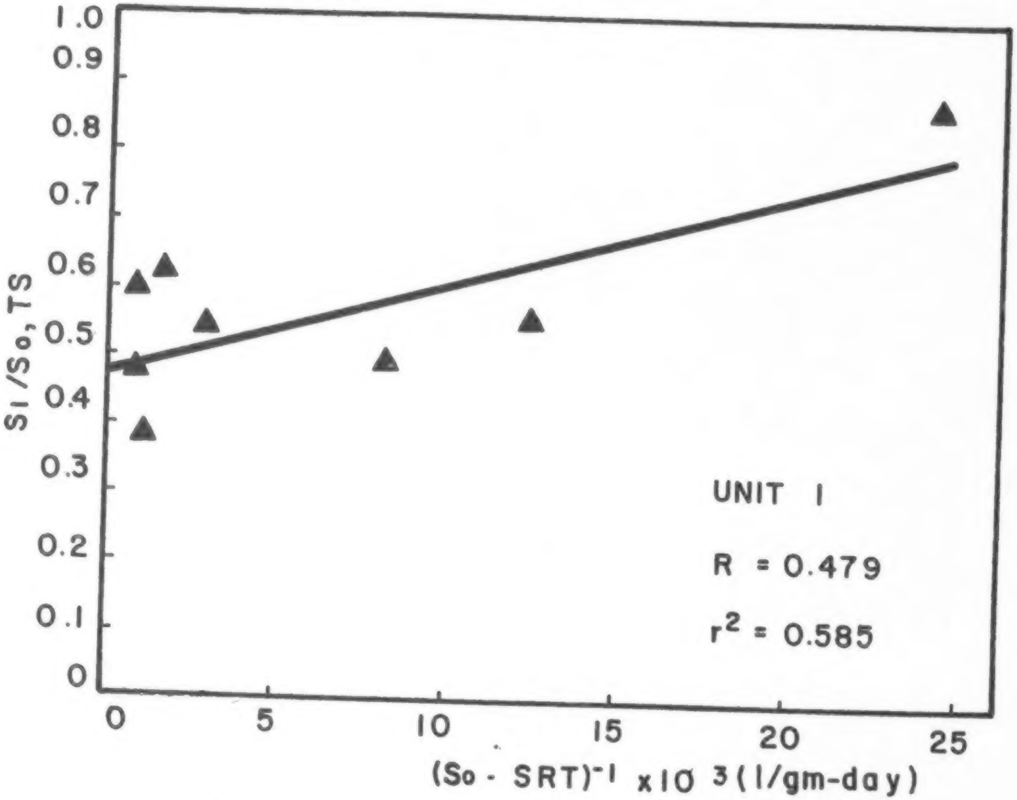

Figure C -41. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 4.

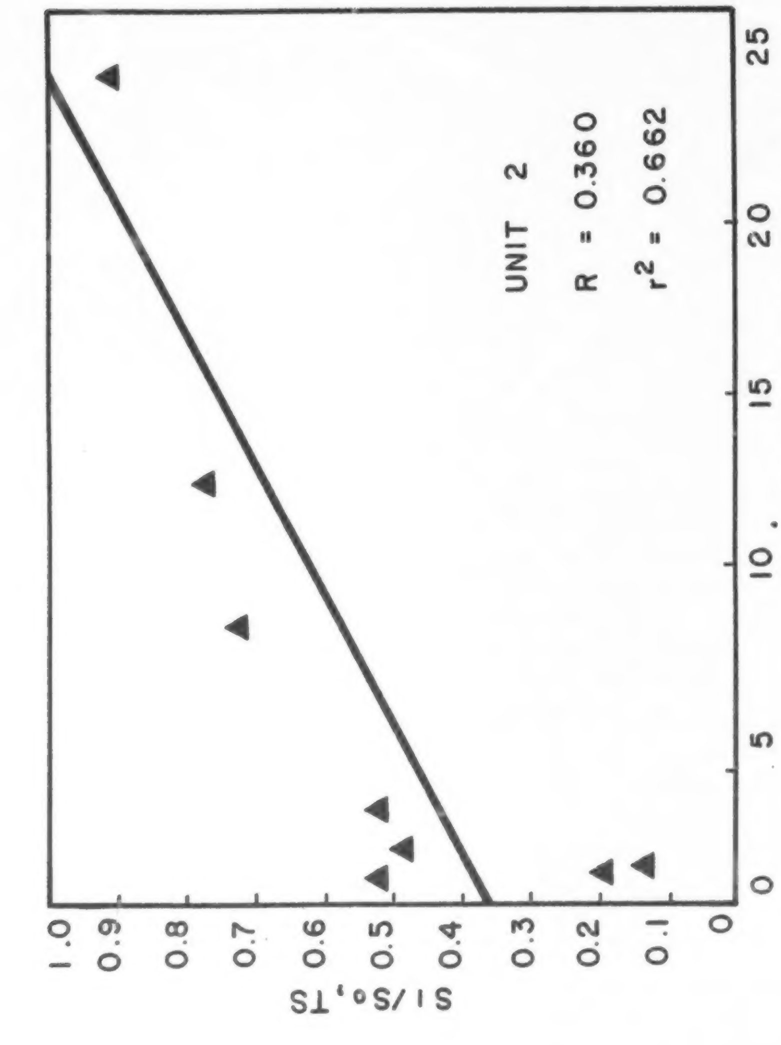

Figure C-42. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 4.

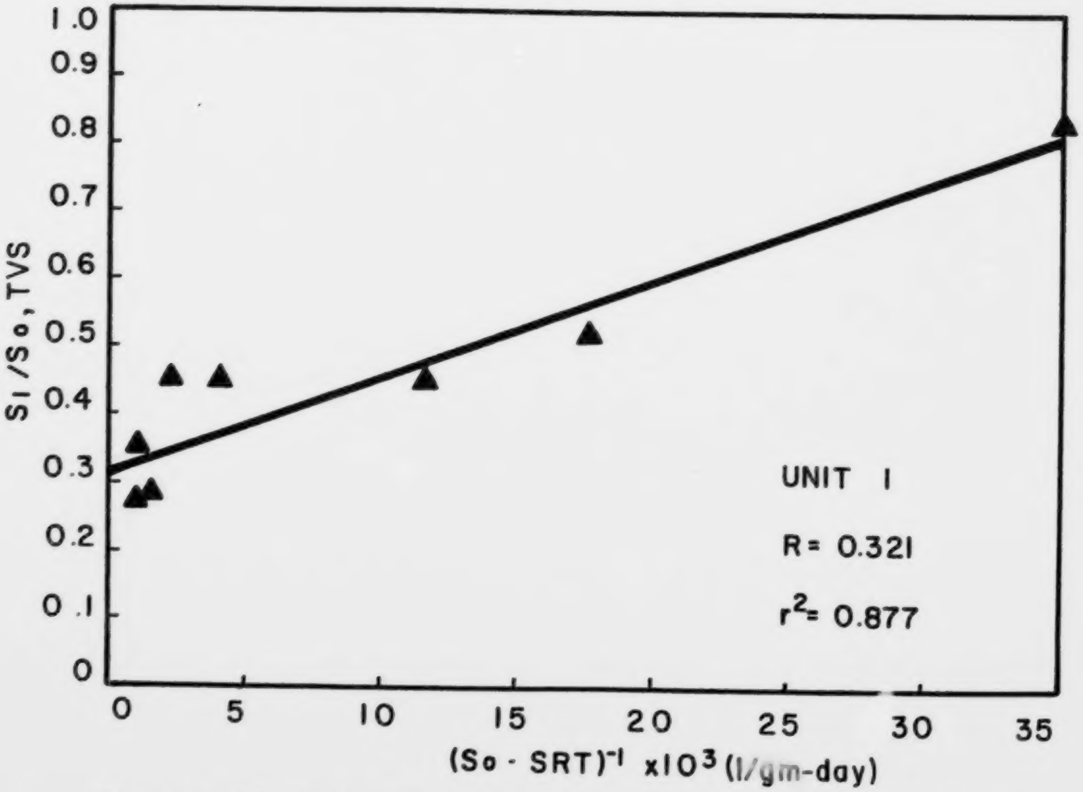

Figure C-43. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 4.

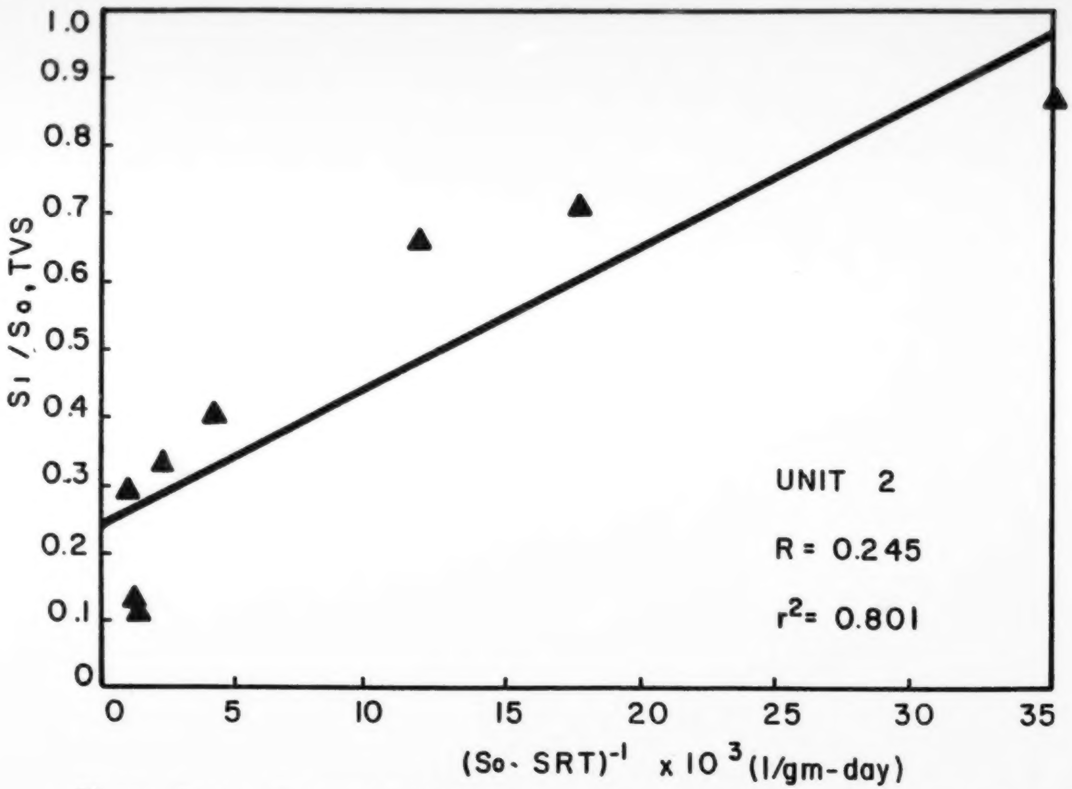

Figure C-44. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 4.

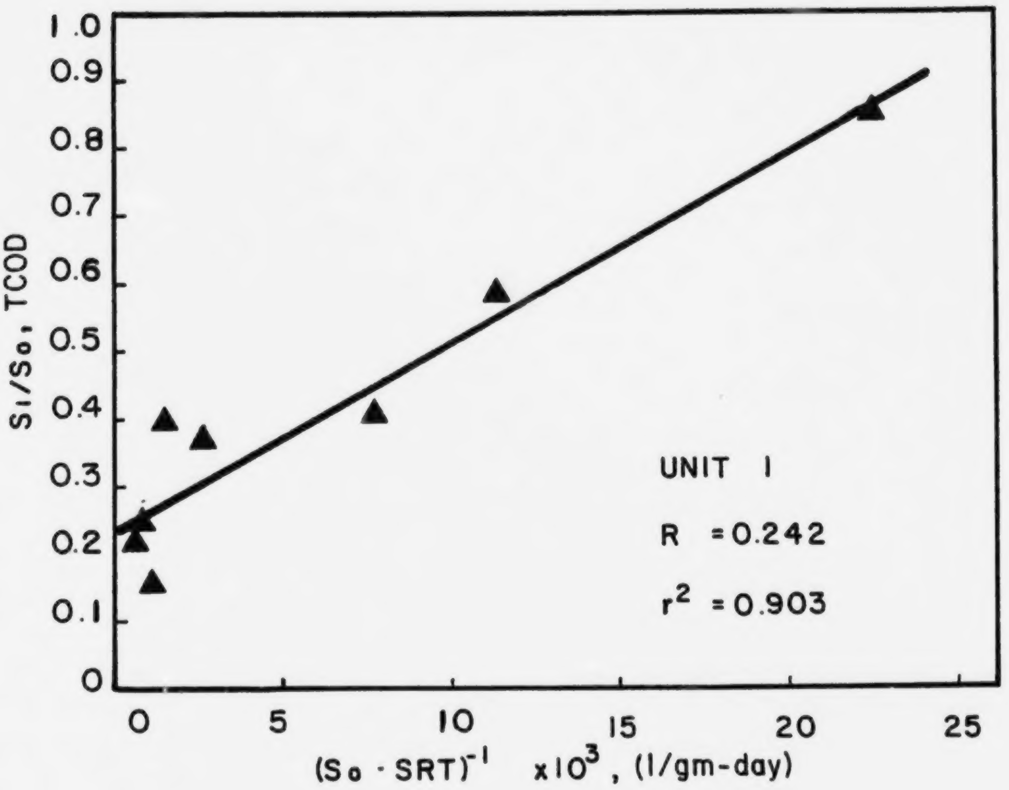

Figure C -45. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 4.

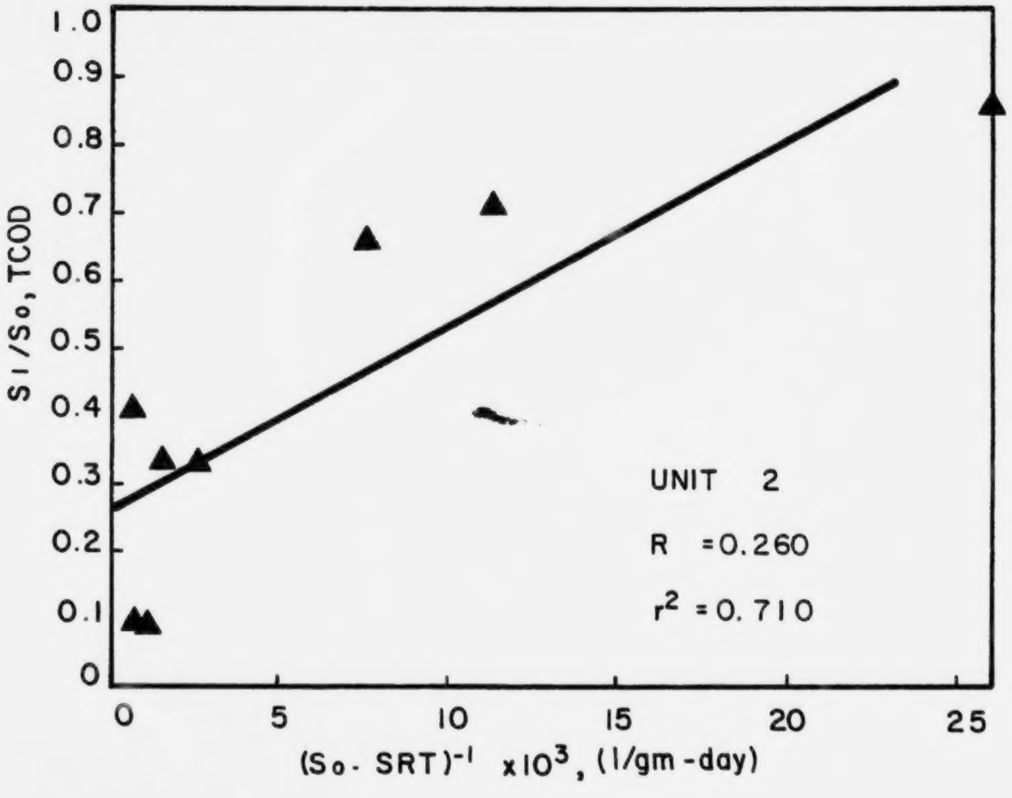

Figure C-46. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 4.

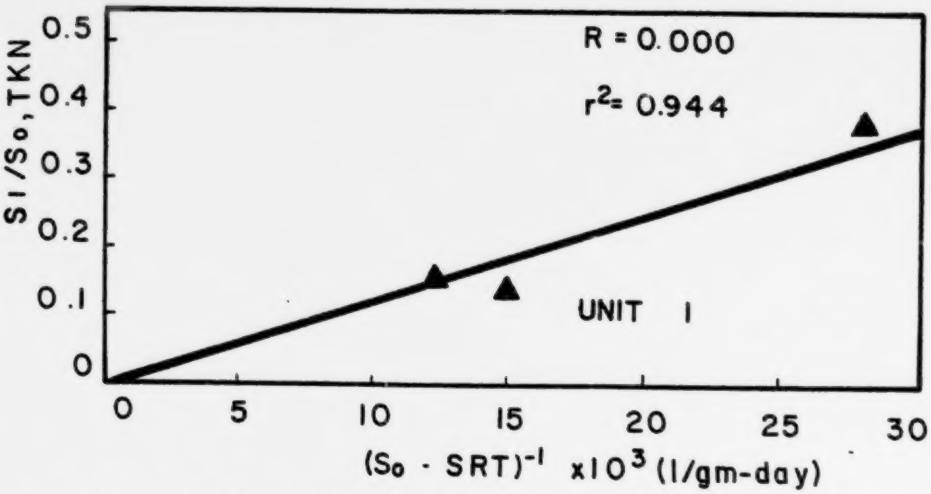

Figure C-47. Determination of the refractory fraction (R) for TKN, biodegradability study no. 4.

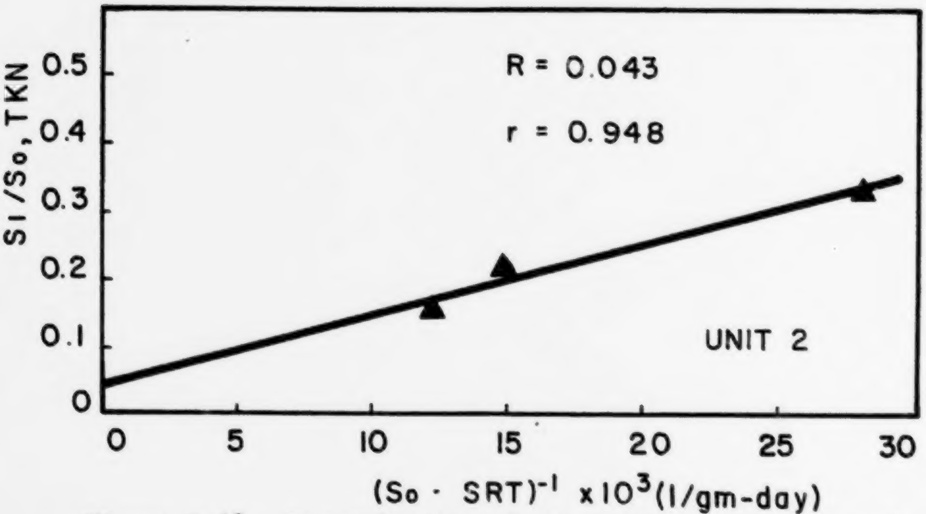

Figure C-48. Determination of the refractory fraction (R) for TKN, biodegradability study no. 4.

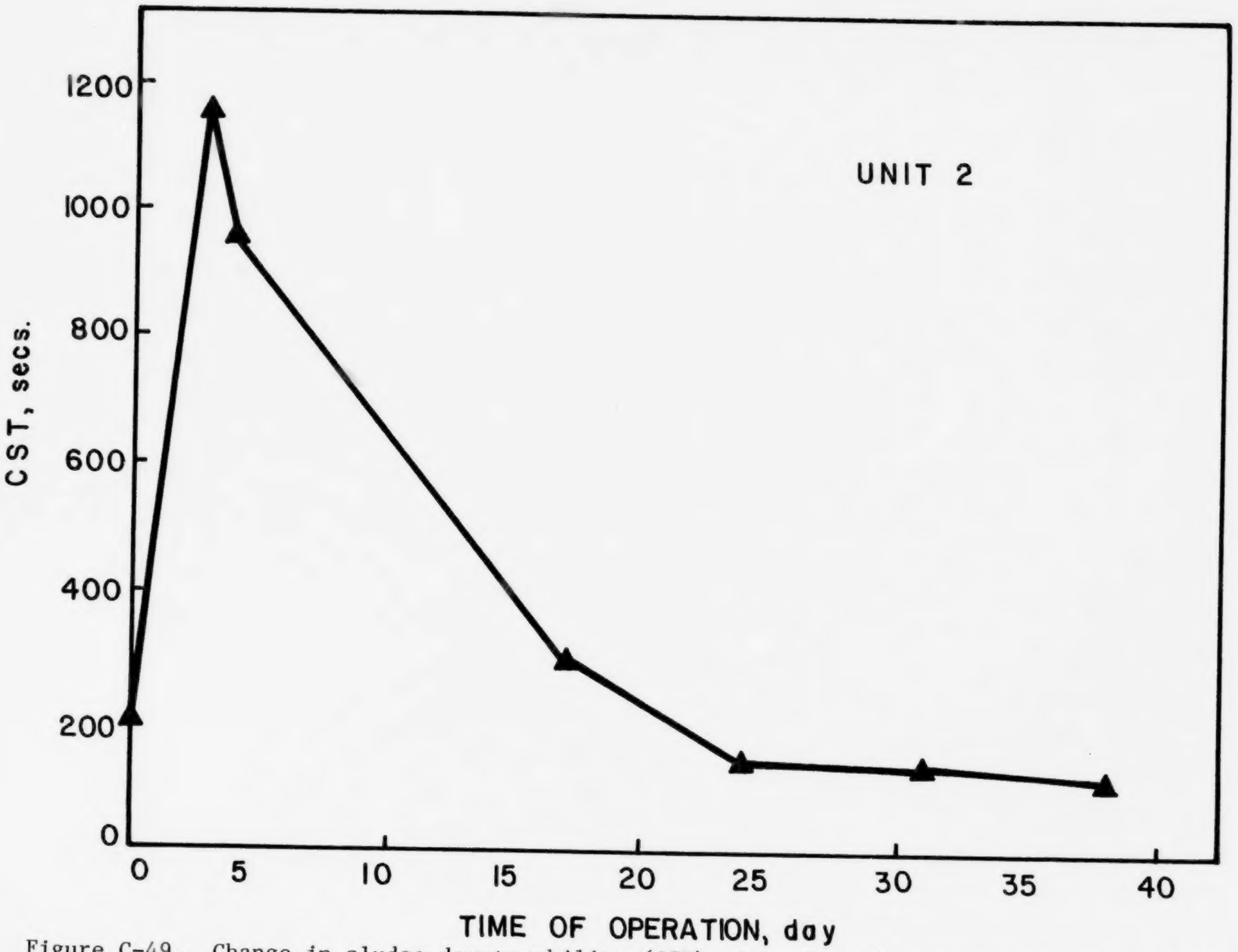

Figure C-49. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 4.

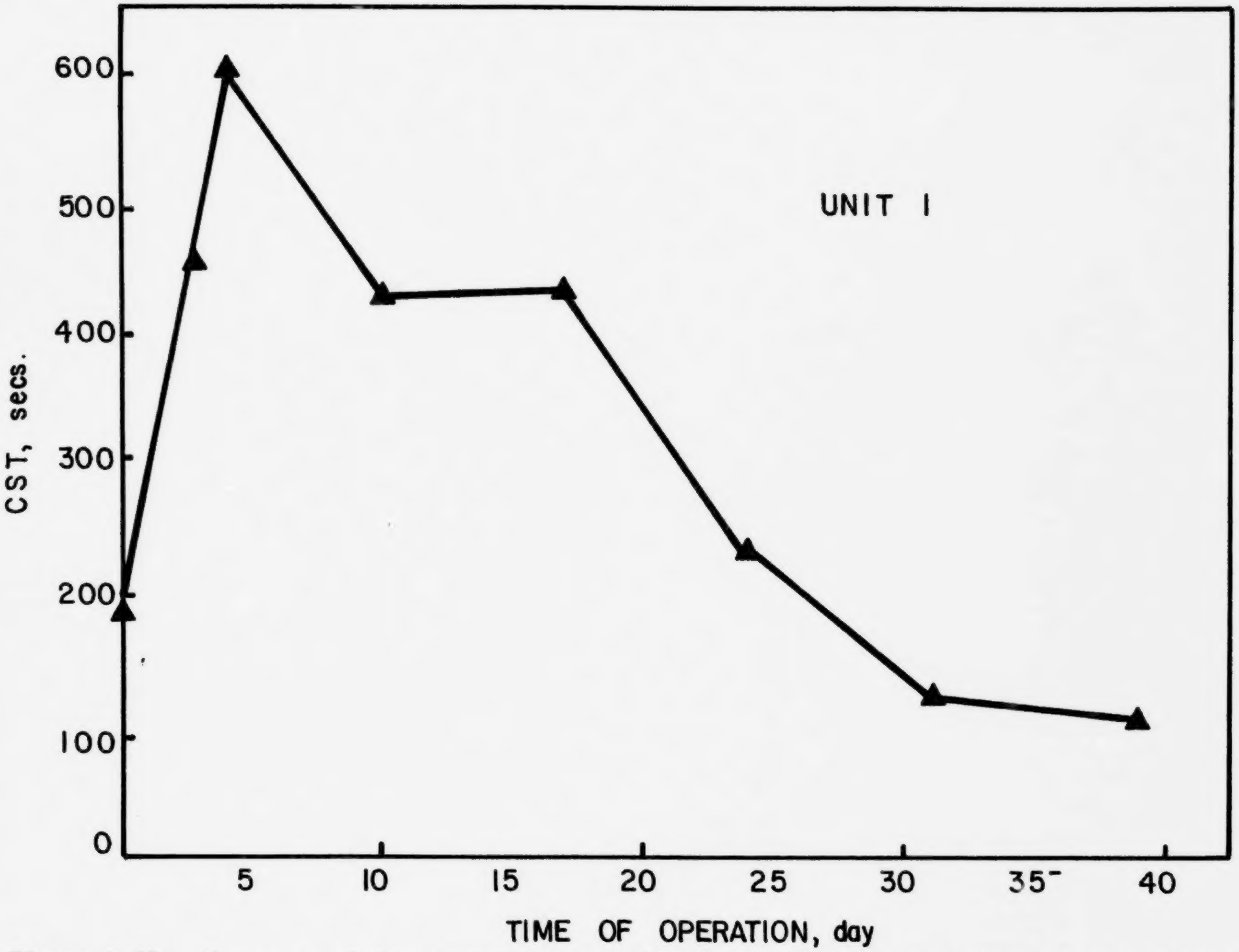

Figure C-50. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 4.

TABLE C-9. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 5, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % $\frac{TVS}{TS}$ |
|---|---|---|---|---|---|---|---|
| 0 | 50.866 | 46.478 | 29.730 | - - | -- | - - | - - |
| 2 | 48.913 | 43.183 | 27.806 | 3.84 | 7.09 | 6.47 | 64.4 |
| 4 | 45.575 | 45.320 | 27.562 | 10.40 | 2.49 | 7.29 | 60.8 |
| 9 | 36.382 | 38.408 | 21.553 | 28.48 | 17.36 | 27.5 | 56.1 |
| 16 | 22.945 | 34.700 | 18.135 | 54.89 | 25.34 | 39.0 | 52.2 |
| 23 | 25.008 | 33.382 | 16.753 | 50.84 | 28.18 | 43.65 | 53.75 |
| 30 | 30.617 | 41.472 | 20.369 | 39.81 | 56.17 | 31.49 | 49.1 |
| 37 | 16.373 | 26.737 | 11.684 | 67.81 | 42.47 | 60.7 | 43.6 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 50.71 | 43-60 |
| D.O. mg/ℓ | 4.18 | .06-7.5 |
| pH | 7.36 | 6.8 -7.95 |
| Evaporation, mℓ/day | 700 | 400 - 800 |
| Airflow Rate, ℓ/sec | .0679 | .063-.079 |

TABLE C-10. RESULTS OF BIODEGRADABILITY STUDY NO. 5, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % $\frac{TVS}{TS}$ |
|---|---|---|---|---|---|---|---|
| 0 | 50.866 | 46.478 | 29.730 | -- | -- | - - | 63.97 |
| 2 | 48.736 | 42.740 | 27.463 | 4.19 | 8.04 | 7.63 | 64.3 |
| 4 | 45.510 | 43.546 | 26.291 | 10.53 | 6.31 | 11.57 | 60.4 |
| 9 | 44.712 | 44.033 | 25.185 | 12.10 | 5.26 | 15.29 | 57.2 |
| 16 | 34.052 | 43.777 | 23.7614 | 33.06 | 5.81 | 20.08 | 54.9 |
| 23 | 35.063 | 33.871 | 13.004 | 31.07 | 27.12 | 56.26 | 38.2 |
| 30 | 27.905 | 38.215 | 18.634 | 34.14 | 17.78 | 37.32 | 48.7 |
| 37 | 20.633 | 31.493 | 14.620 | 59.44 | 68.54 | 50.84 | 46.4 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature °C | 50.329 | 44-60 |
| D.O. mg/ℓ | 4.02 | .05-8.0 |
| pH | 7.23 | 6.5 -7.85 |
| Evaporation | 450 | 360 - 800 |
| Airflow rate, ℓ/sec | .0679 | .063-.079 |

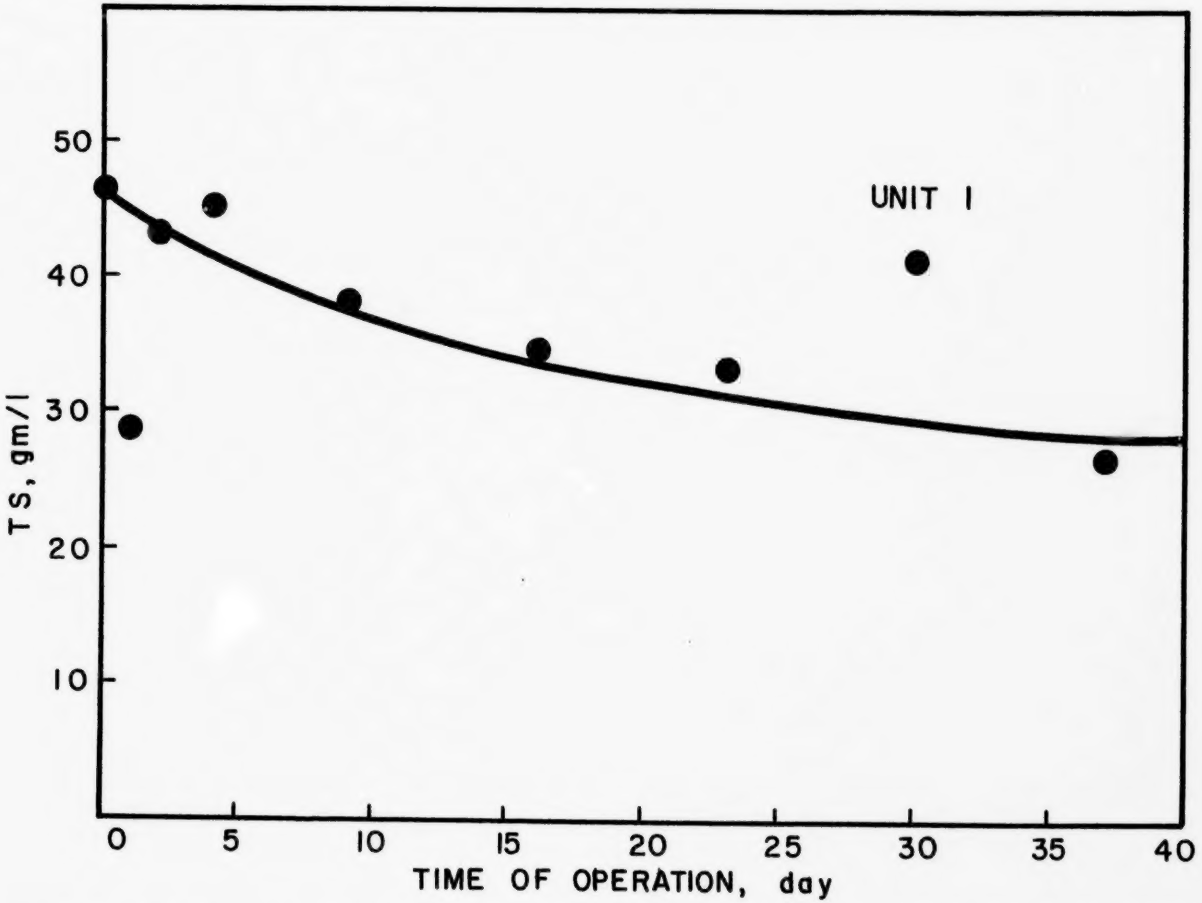

Figure C-51. Total Solids (TS) removal as related to time of operation, Biodegradability Study No. 5.

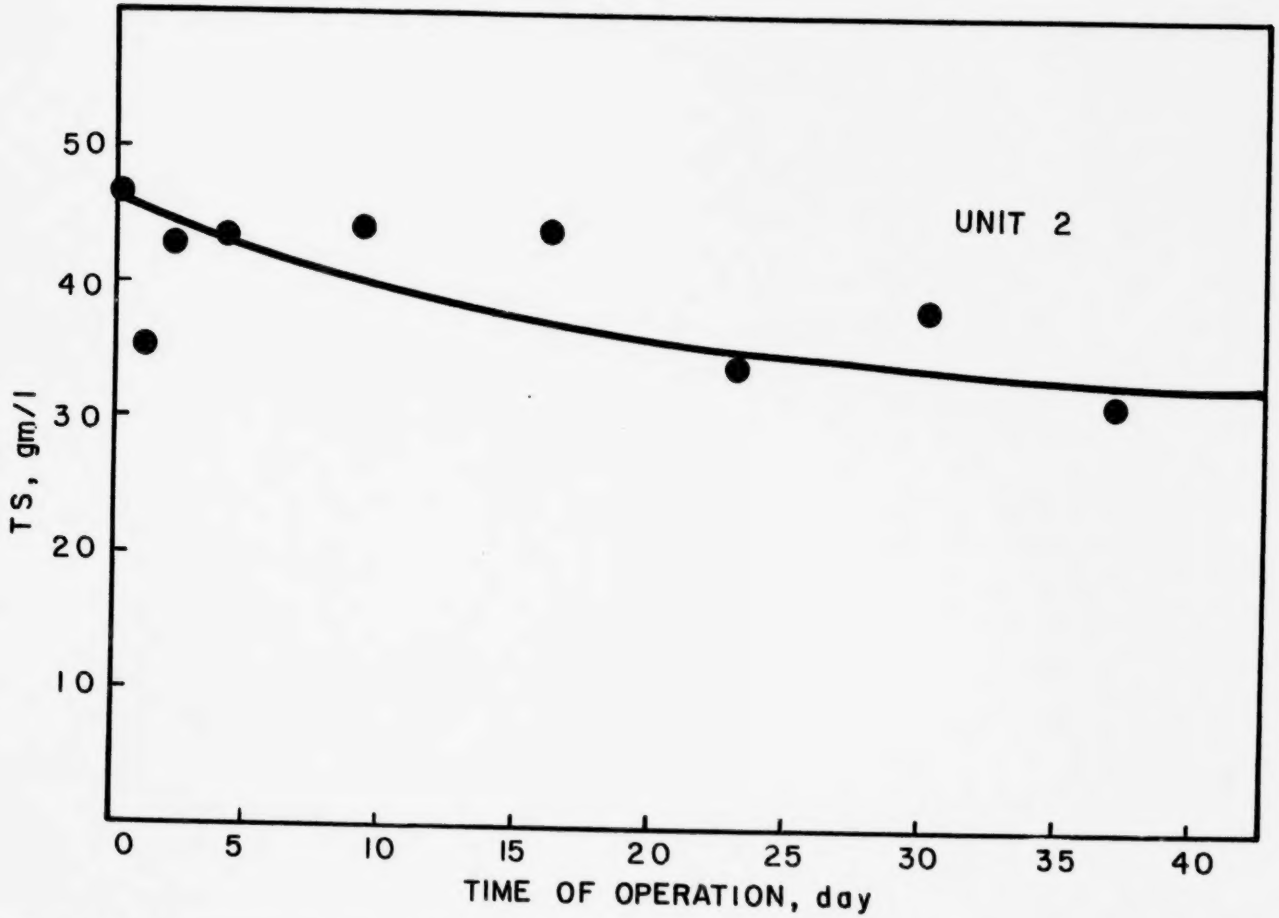

Figure C -52. Total Solids (TS) removal as related to time of operation, Biodegradability Study No. 5.

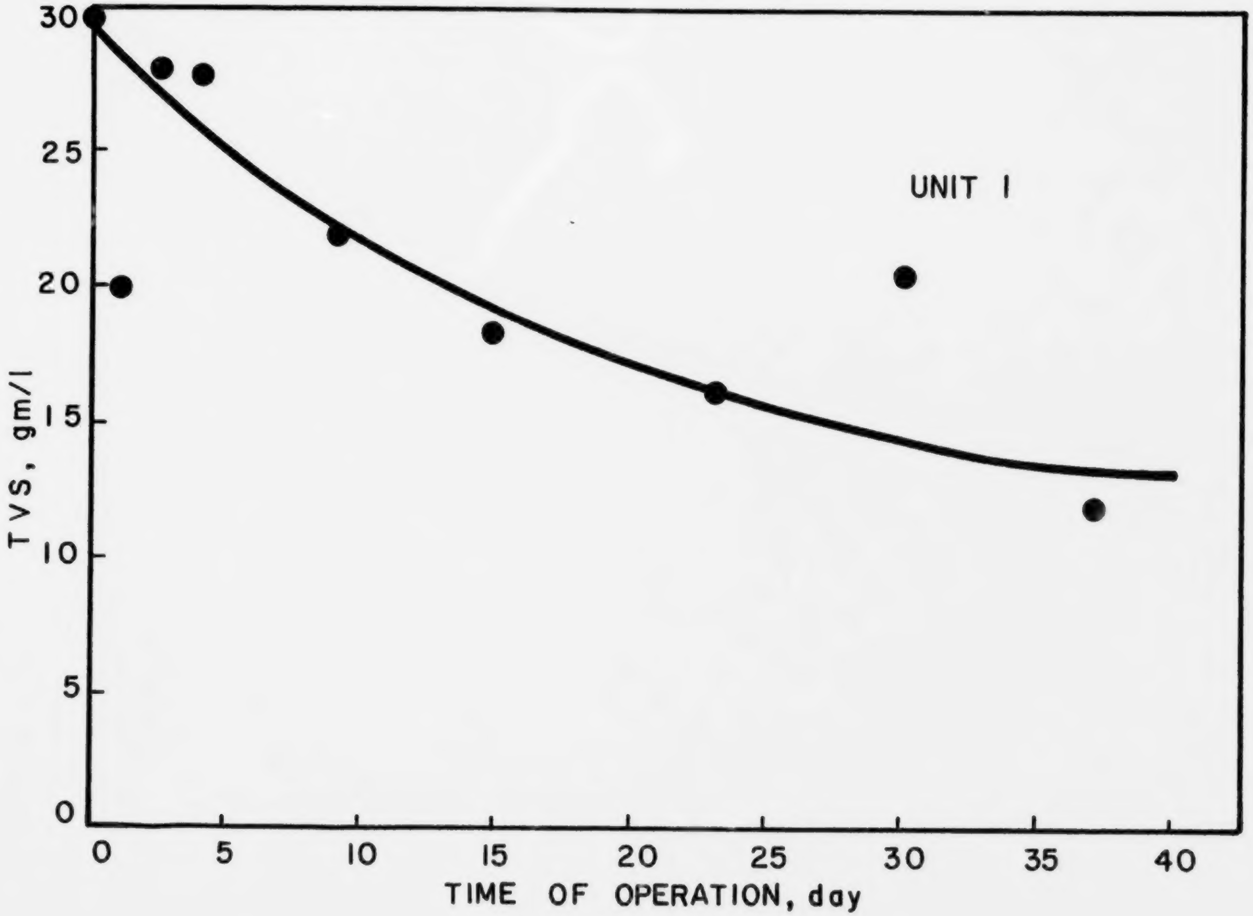

Figure C -53. Total Volatile Solids (TVS) removal as related to time of operation, Biodegradability Study no. 5.

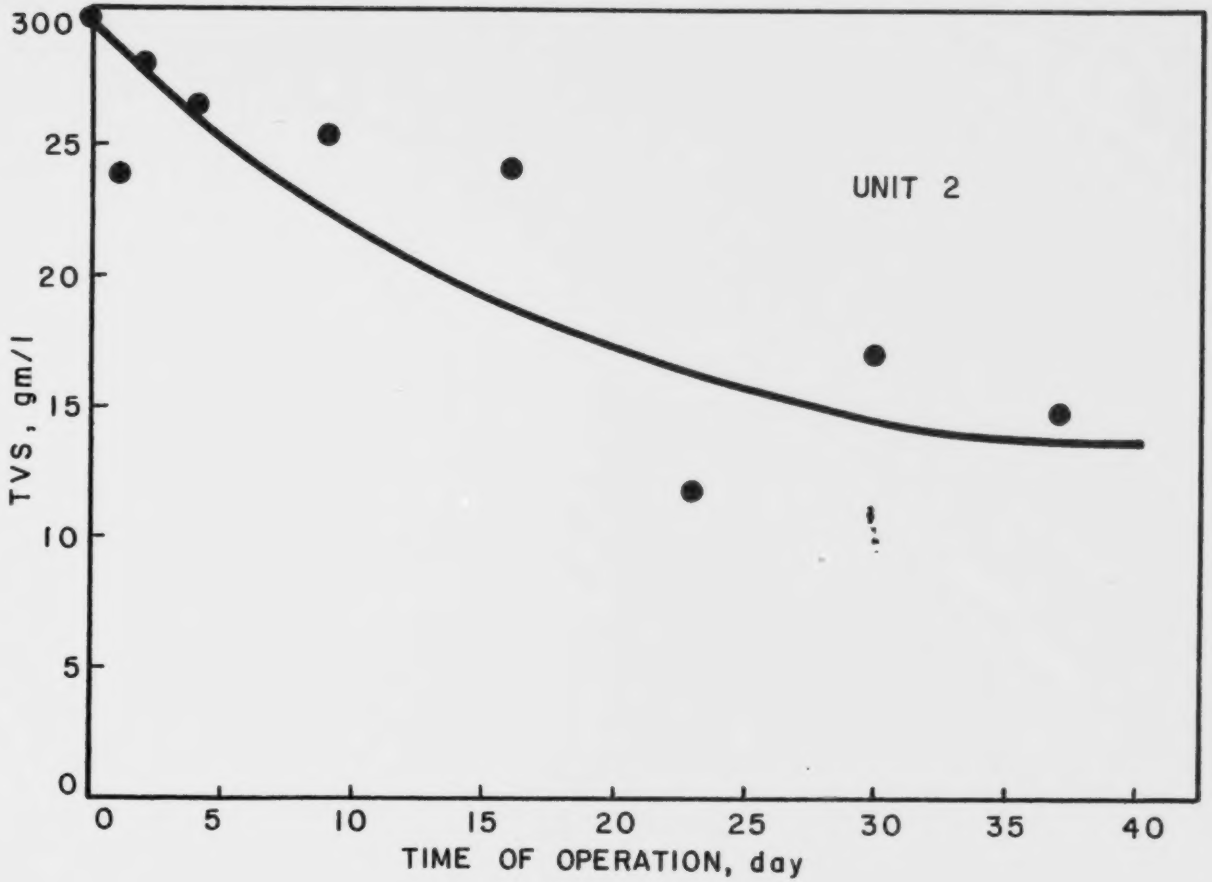

Figure C-54. Total Volatile Solids (TVS) removal as related to time of operation, Biodegradability Study No. 5.

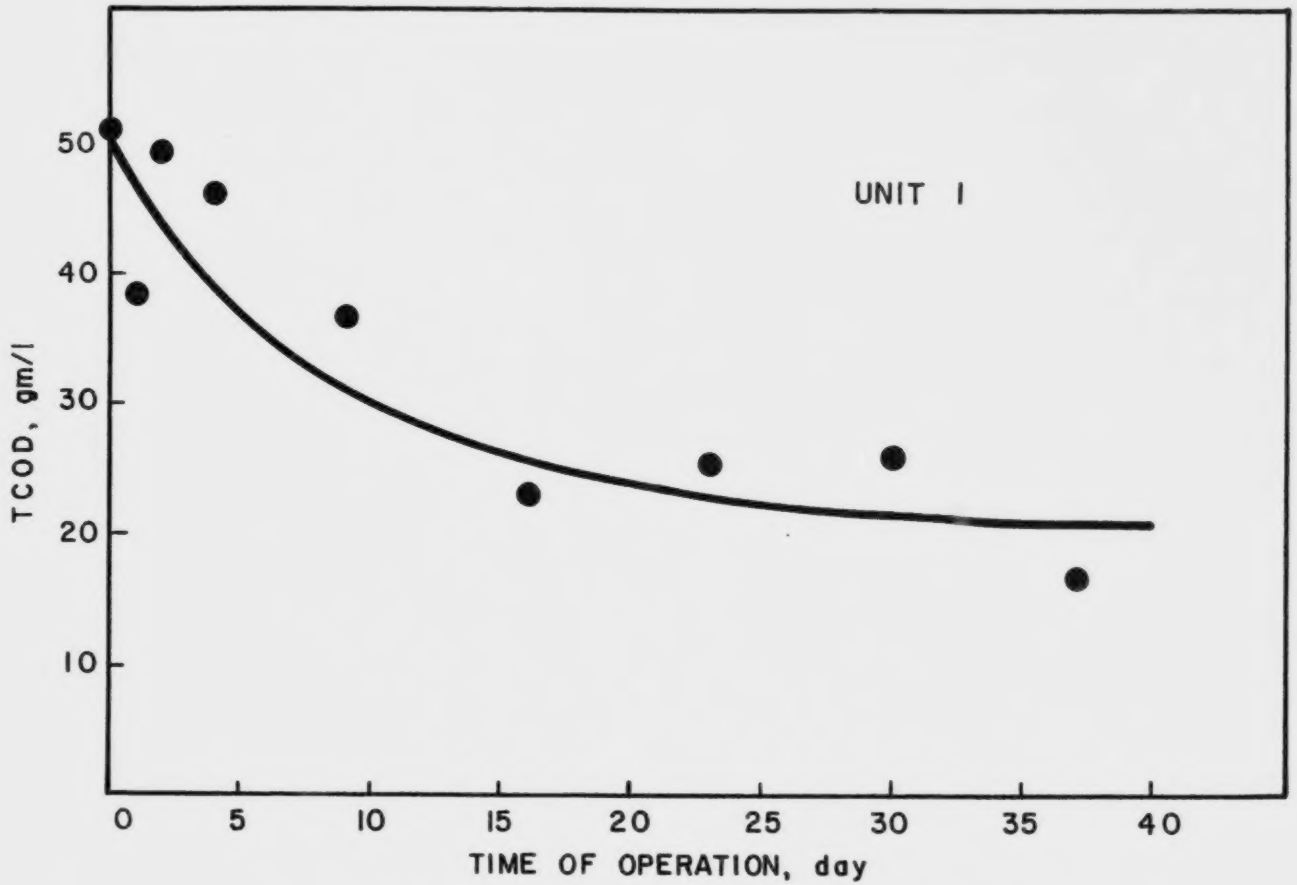

Figure C-55. Total Chemical Oxygen Demand (TCOD) removal as related to time of operation, Biodegradability Study No. 5.

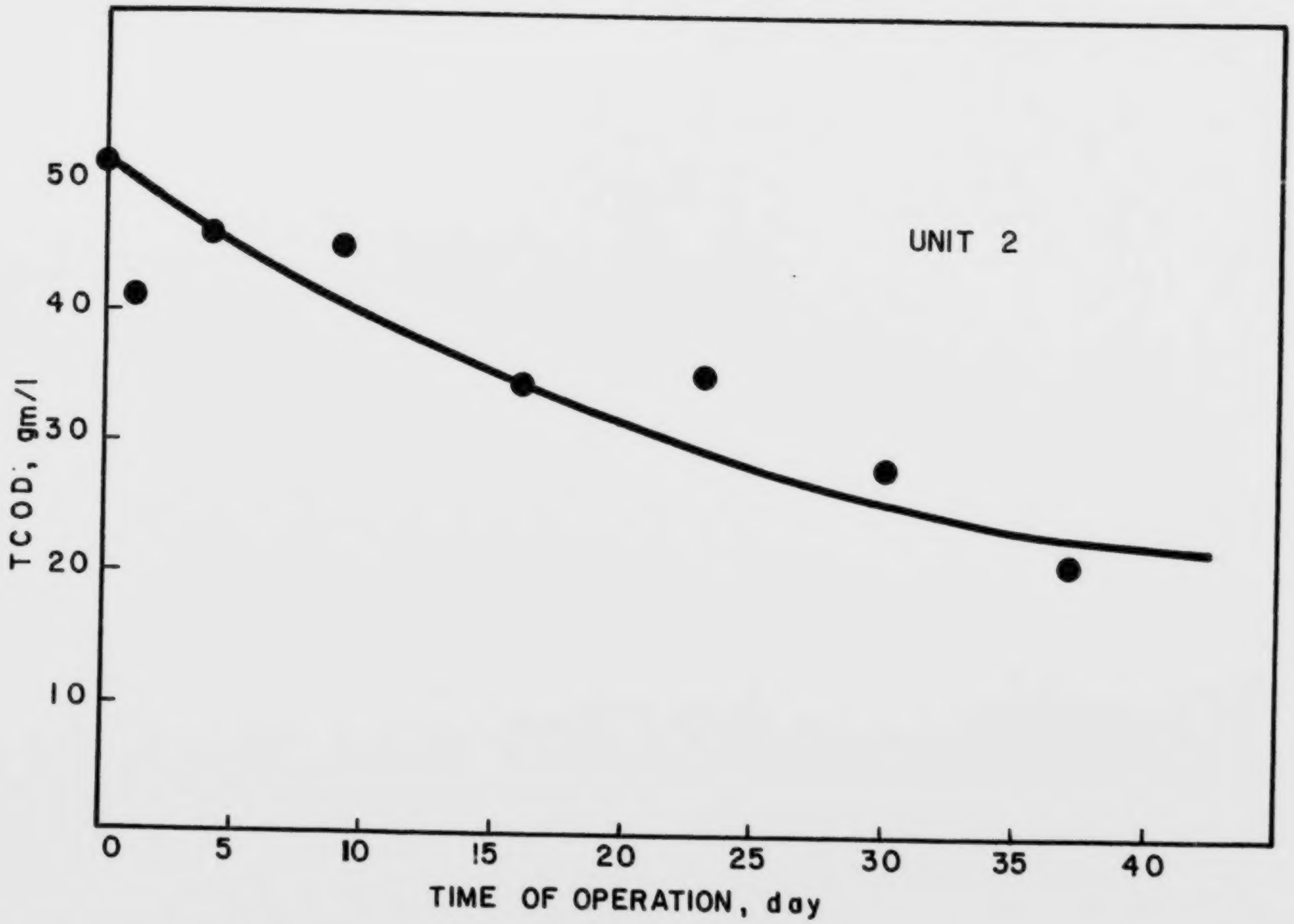

Figure C-56. Total Chemical Oxygen Demand (TCOD) removal as related to time of operation, Biodegradability Study No. 5.

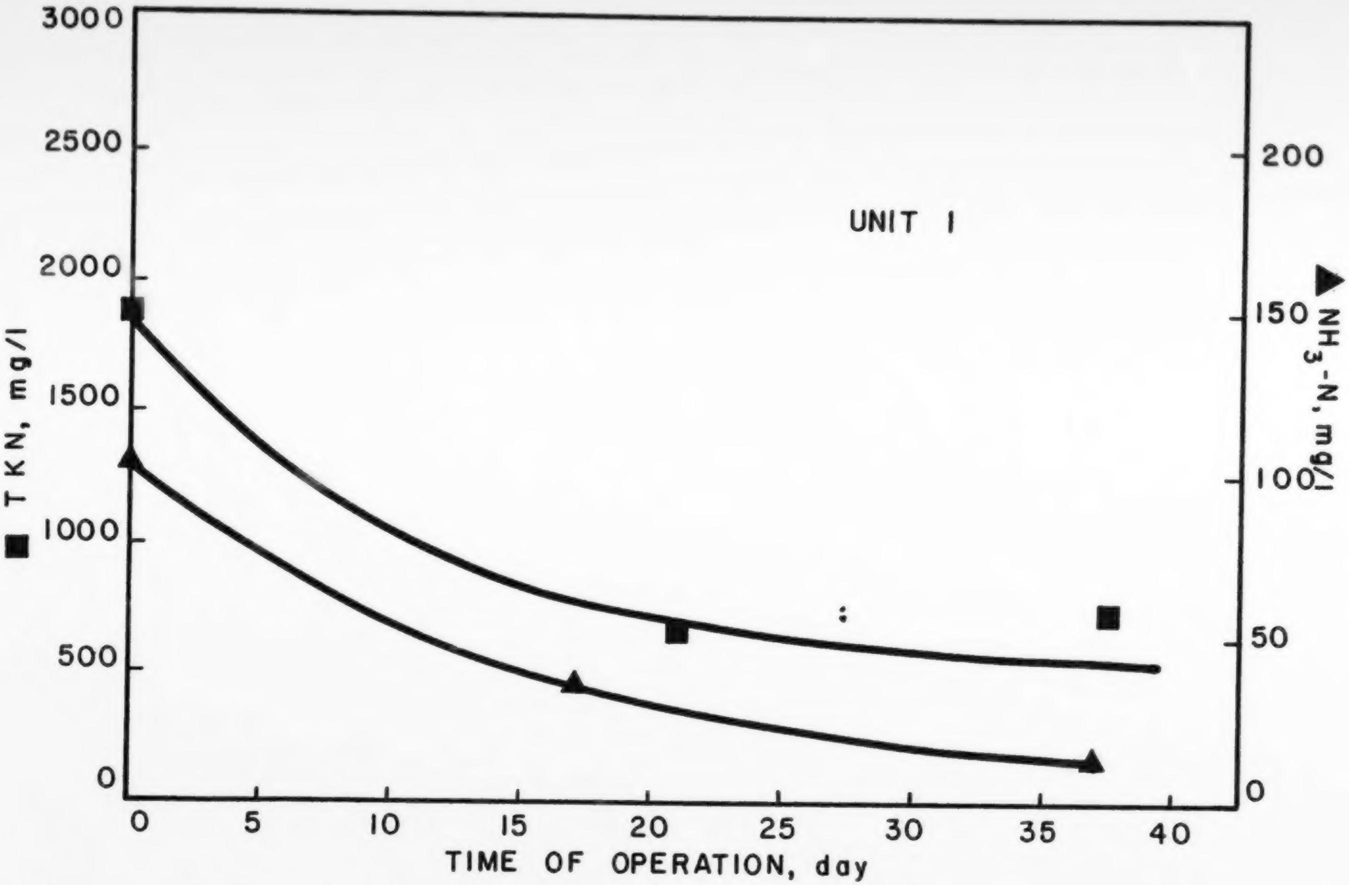

Figure C-57. Total Kjeldahl Nitrogen (TKN) and Ammonia ($NH_3$-N) removal as related to time of operation, Biodegradability Study No. 5.

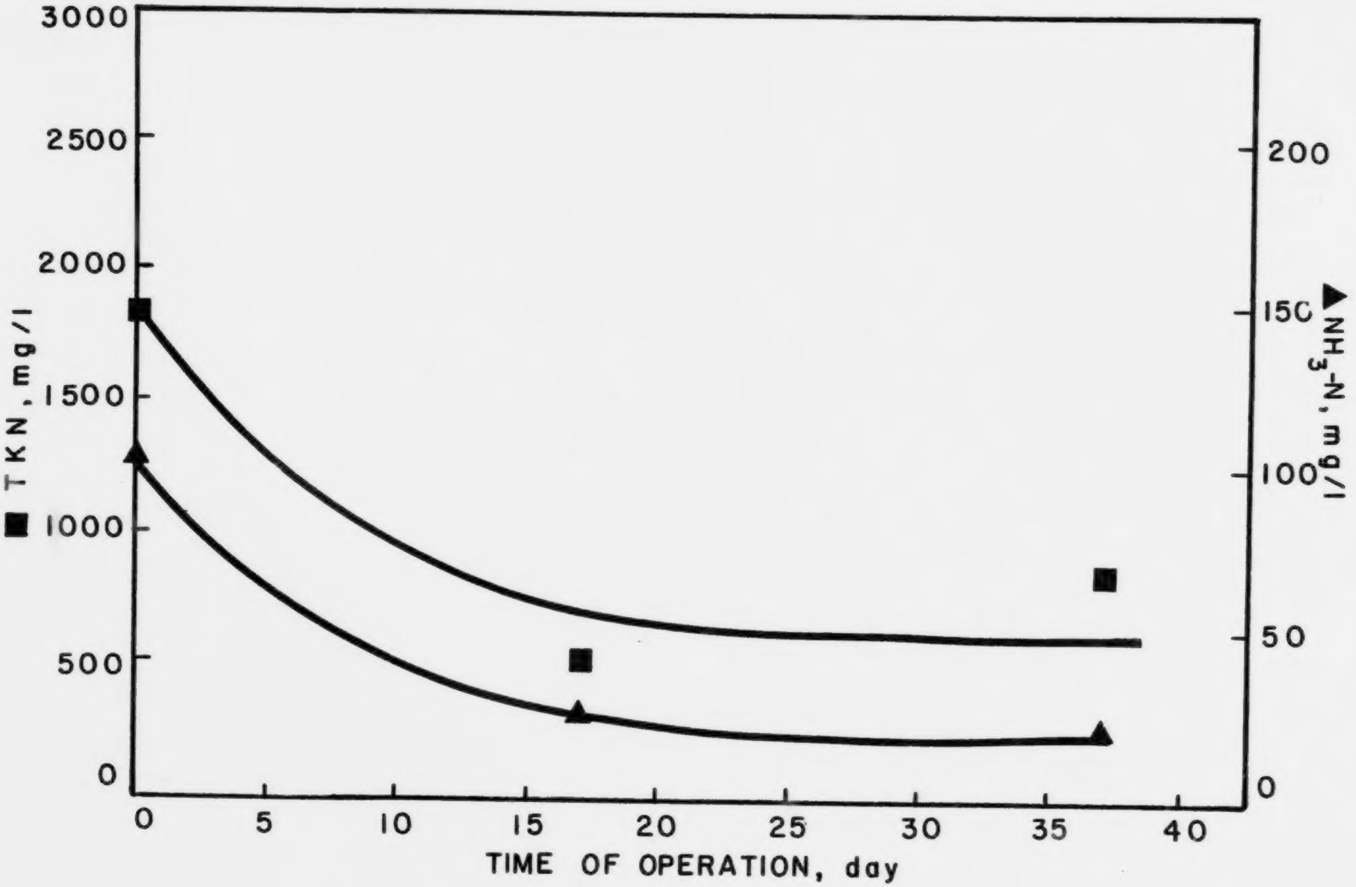

Figure G-58. Total Kjeldahl Nitrogen (TKN) and Ammonia ($NH_3$-N) removal as related to time of operation, Biodegradability Study No. 5.

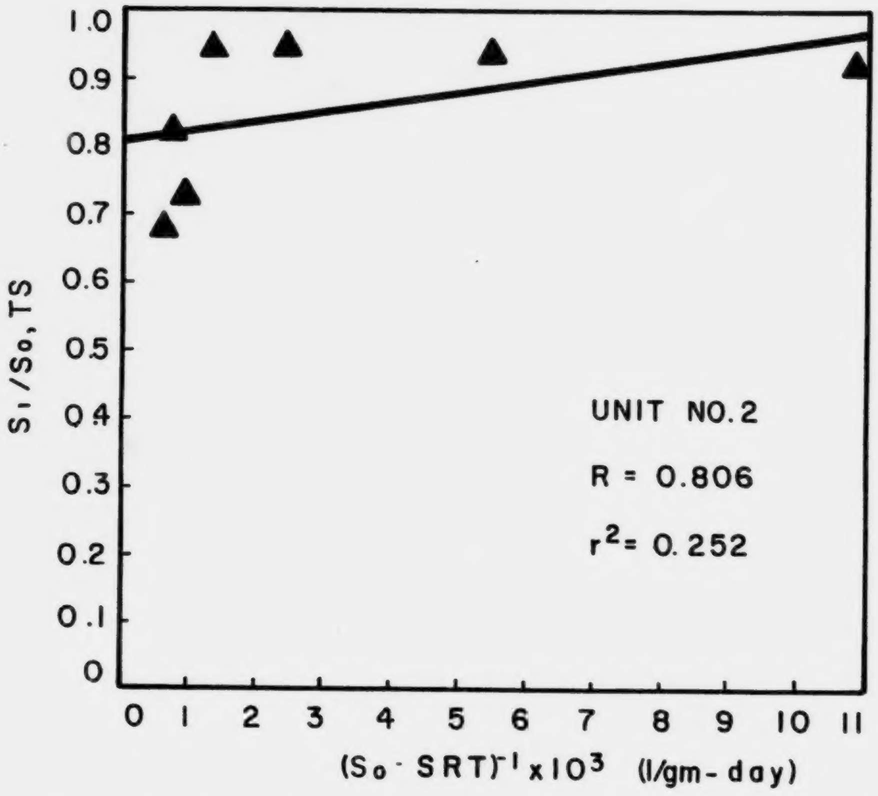

Figure C-59. Determination of the Refractory Fraction (R) for Total Solids (TS), Biodegradability Study No. 5.

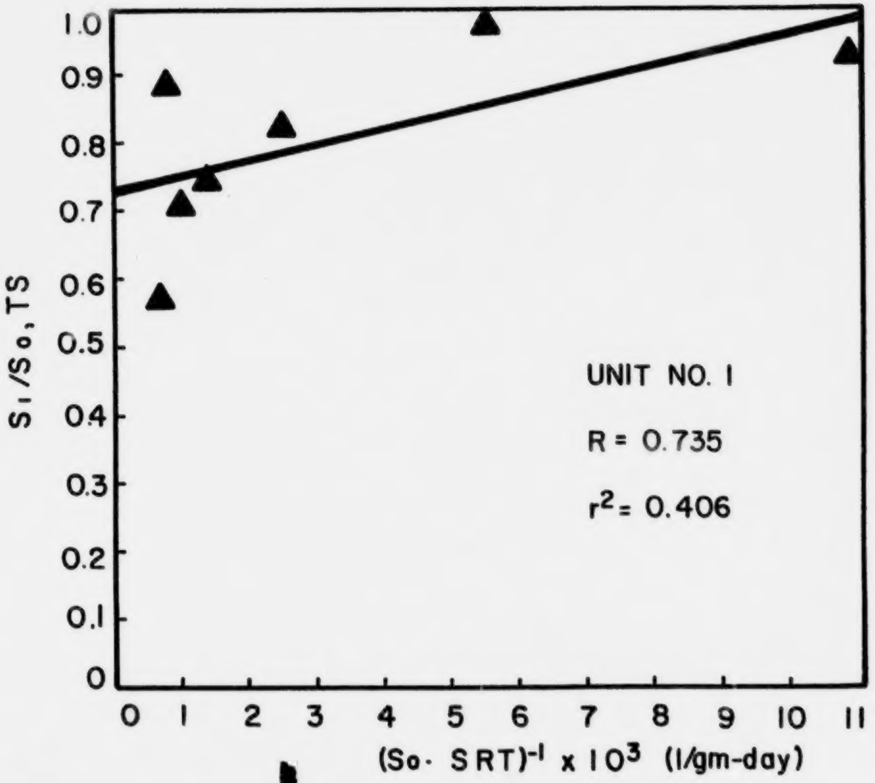

Figure G-60. Determination of the Refractory Fraction (R) for Total Solids (TS), Biodegradability Study No. 5.

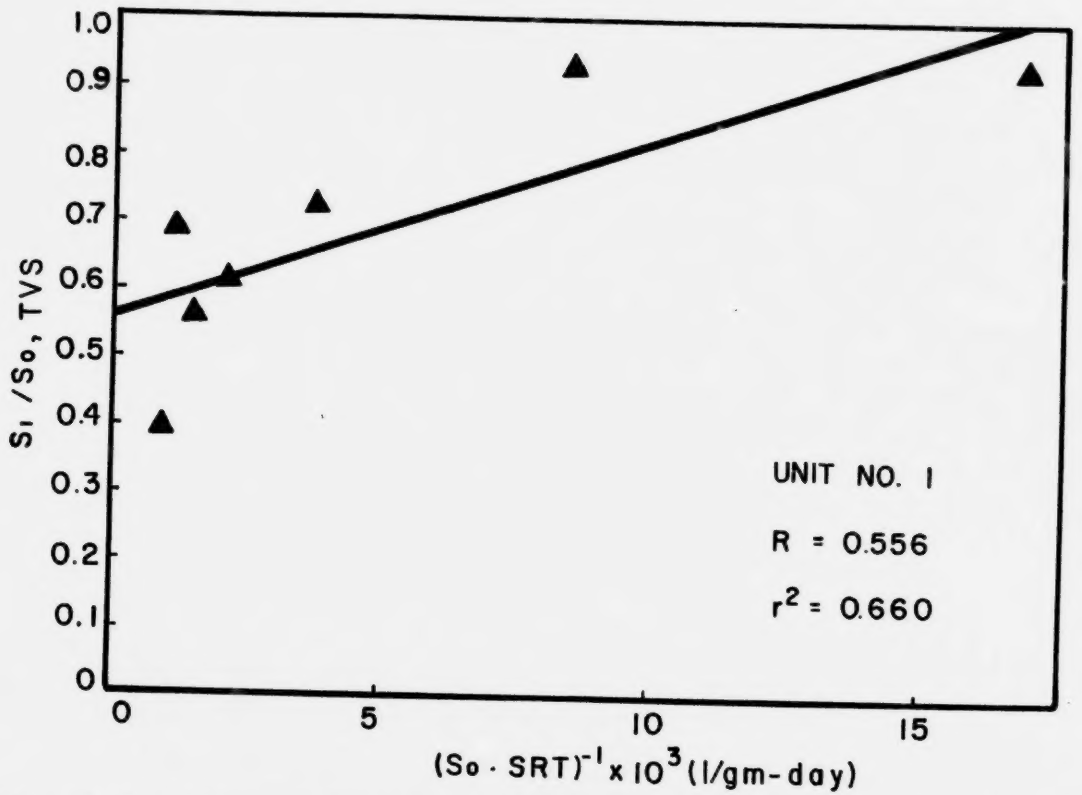

Figure C-61. Determination of the Refractory Fraction (R) for Total Volatile Solids (TVS), Biodegradability Study No. 5.

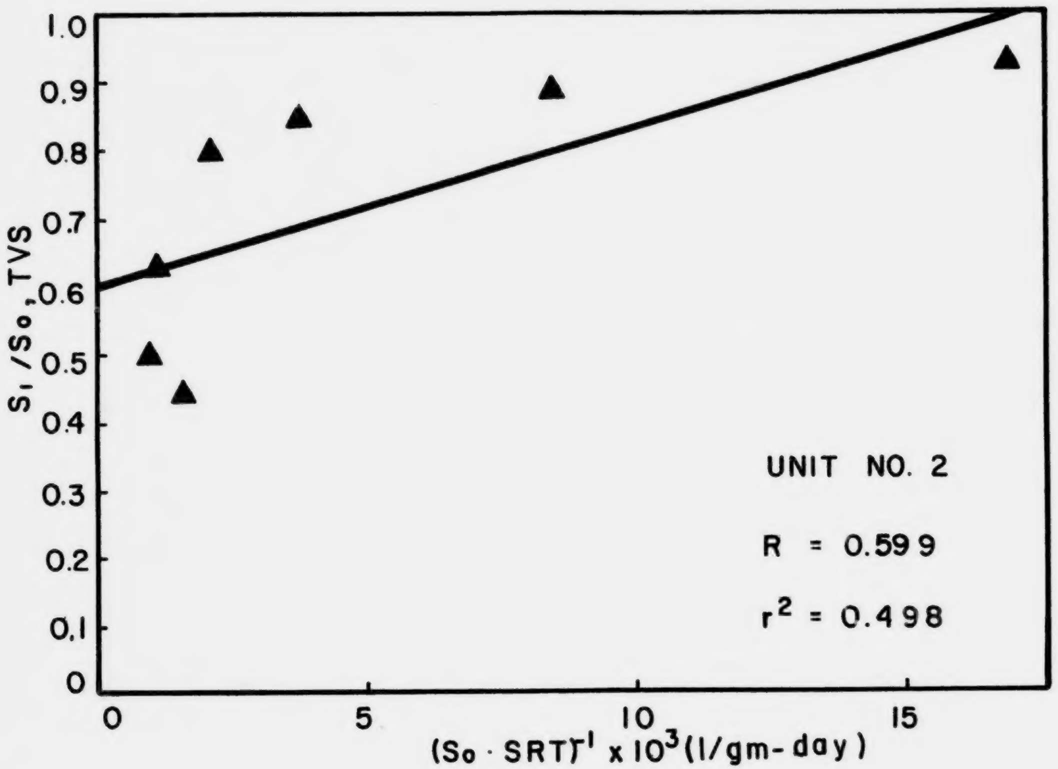

Figure C-62. Determination of the Refractory Fraction (R) for Total Volatile Solids (TVS), Biodegradability Study No. 5.

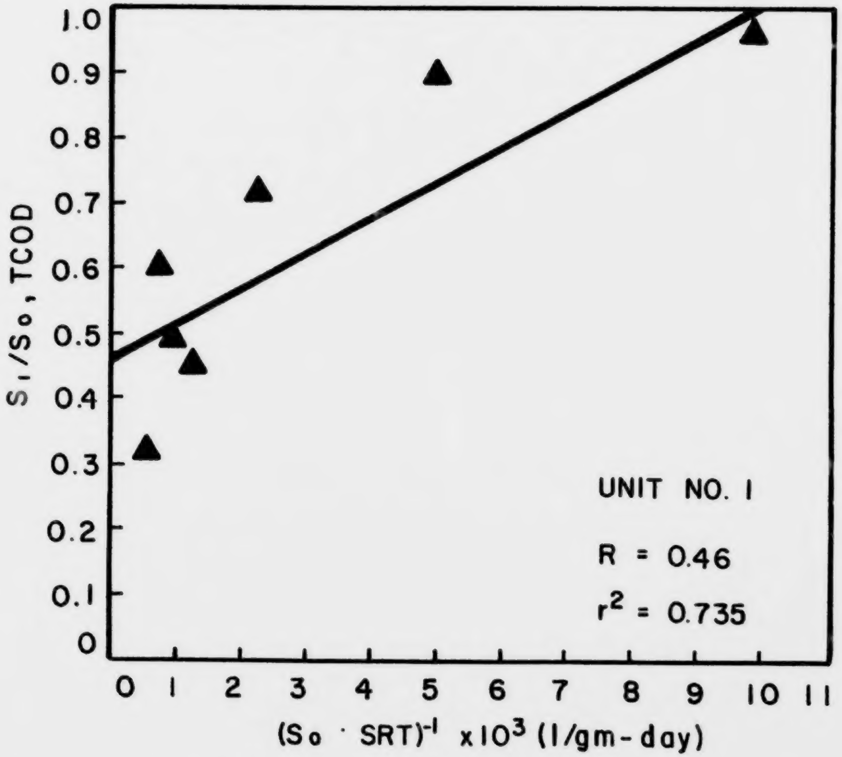

Figure C-63. Determination of the Refractory Fraction (R) for Total Chemical Oxygen Demand (TCOD), Biodegradability Study No. 5.

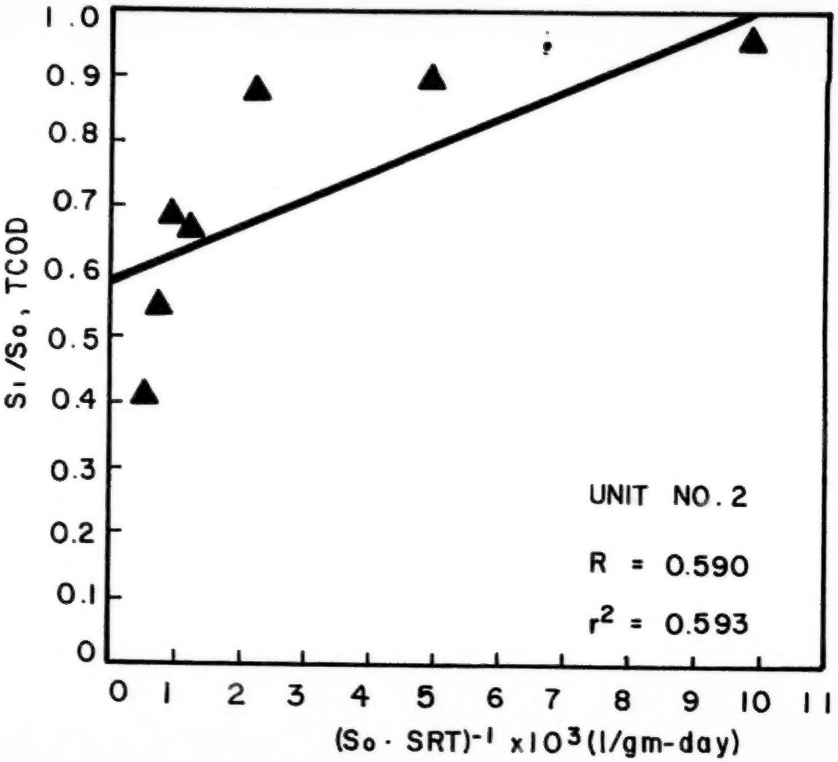

Figure C-64. Determination of the Refractory Fraction (R) for Total Chemical Oxygen Demand (TCOD), Biodegradability Study No. 5.

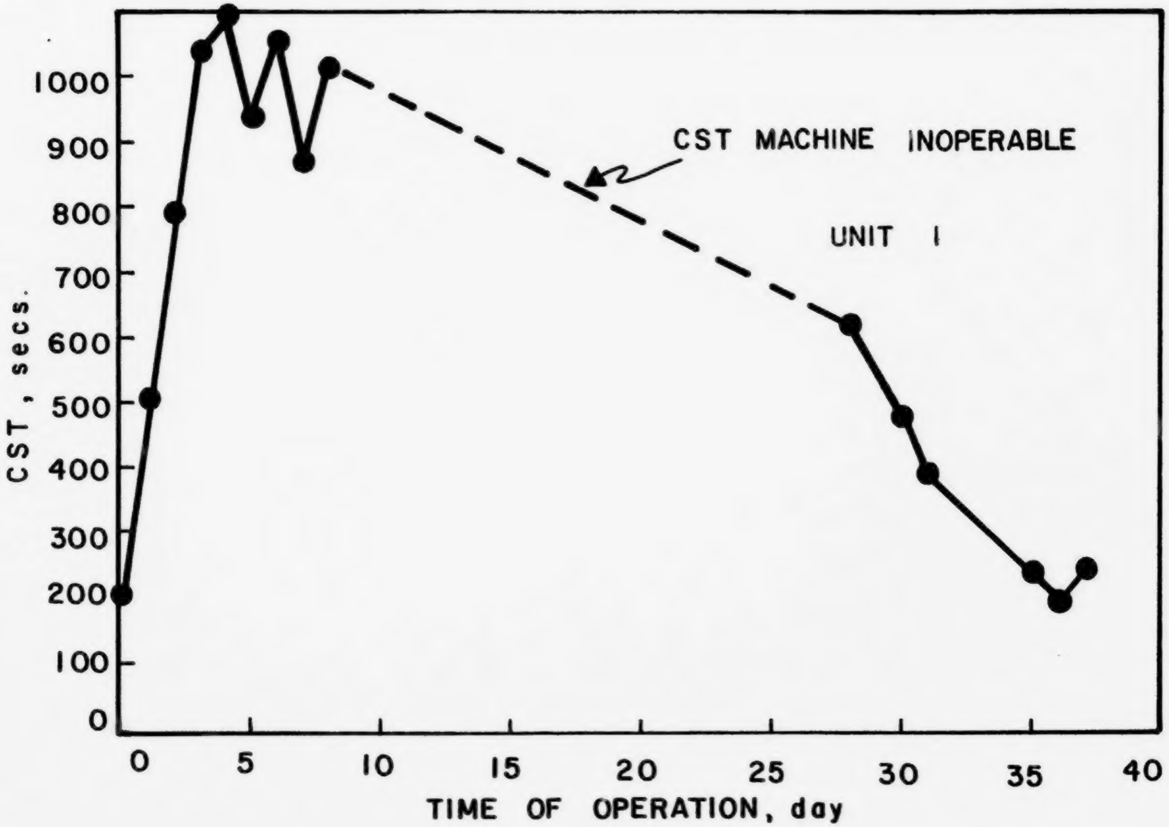

Figure C-65. Change in Sludge Dewaterability (CST) with time of operation, Biodegradability Study No. 5.

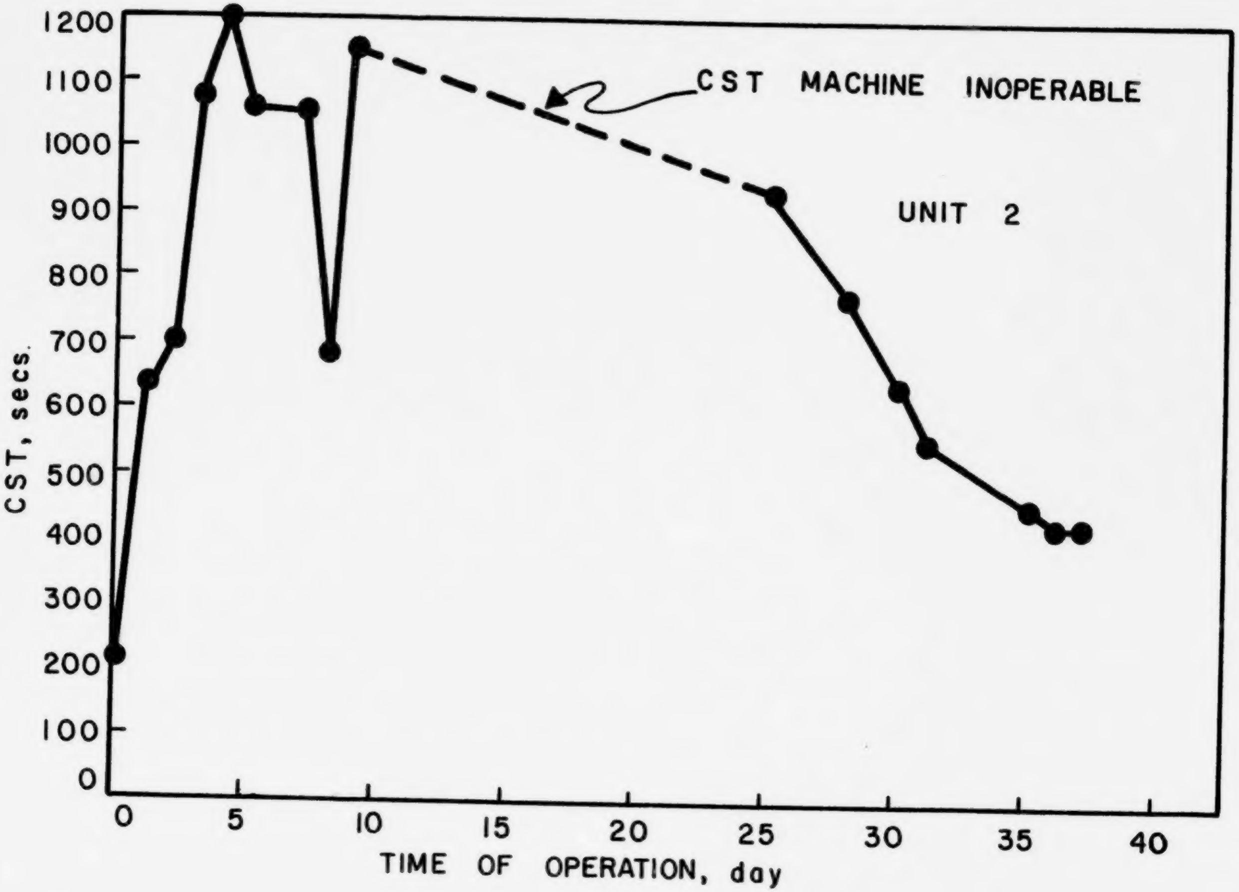

Figure C -66. Change in Sludge Dewaterability (CST) with time of operation, Biodegradability Study No. 5.

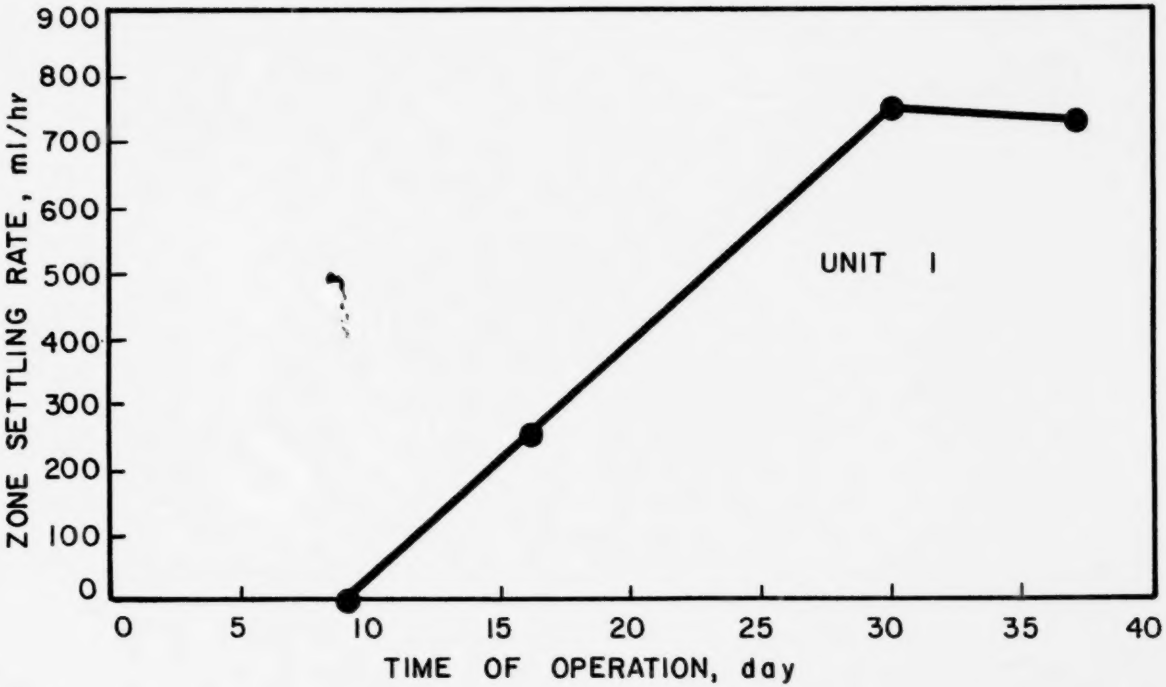

Figure C-67. Zone Settling Rate (ZSR) as related to time of operation, Biodegradability Study No. 5.

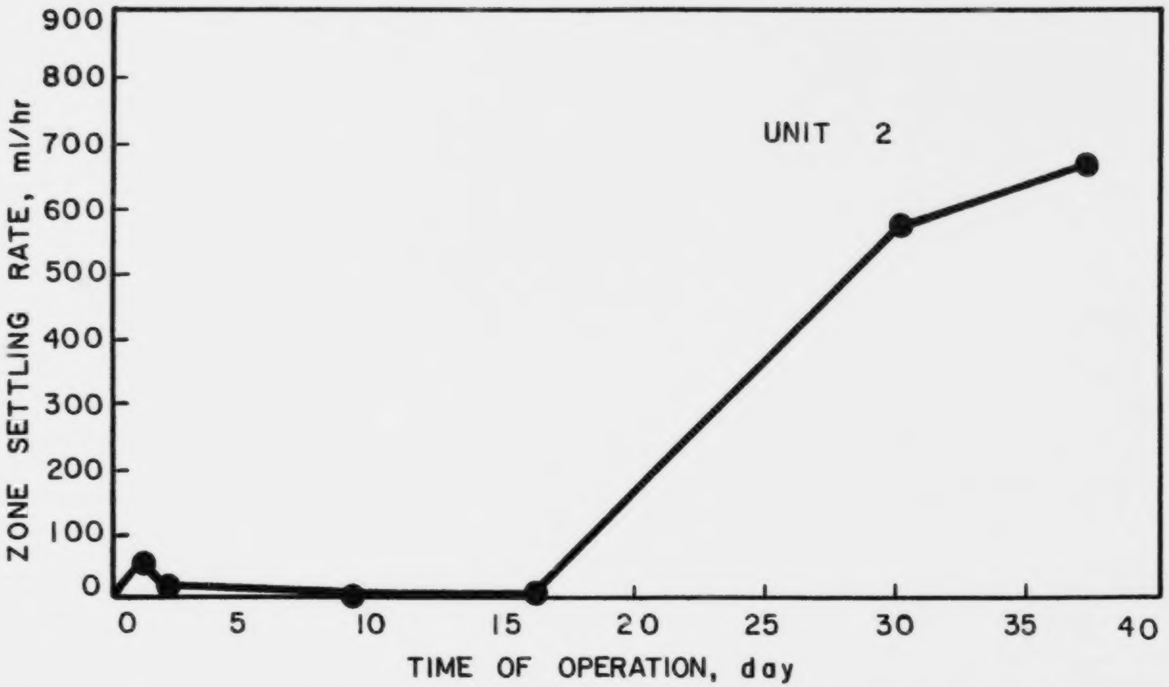

Figure C-68. Zone Settling Rate (ZSR) as related to time of operation, Biodegradability Study No. 5.

TABLE C-11. RESULTS OF BIODEGRADABILITY STUDY NO. 6, UNIT NO. 3

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal | | | % $\frac{TVS}{TS}$ |
|---|---|---|---|---|---|---|---|
| | | | | TCOD | TS | TVS | |
| 0 | 40.645 | 35.050 | 23.149 | - - | - - | - - | 65.9 |
| 2 | 37.221 | 33.310 | 21.724 | 8.42 | 4.97 | 6.15 | 65.2 |
| 5 | 27.476 | 27.494 | 16.435 | 32.40 | 21.56 | 29.0 | 58.1 |
| 12 | 23.115 | 21.980 | 13.716 | 43.13 | 28.73 | 40.75 | 54.9 |
| 19 | 16.535 | 21.584 | 11.357 | 59.32 | 38.42 | 50.94 | 52.6 |
| 27 | 16.782 | 21.052 | 10.349 | 58.71 | 39.94 | 55.29 | 49.2 |
| 31 | 0.941 | 1.891 | 1.486 | 97.70 | 94.60 | 93.58 | 78.6 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 52.61 | 44-54 |
| D.O., mg/ℓ | 4.71 | 0.1-6.5 |
| pH | 7.48 | 7.2-7.8 |
| Evaporation, mℓ/day | 500 | 400-700 |
| Airflow rate, ℓ/sec | .0657 | .047-.079 |

TABLE C-12. RESULTS OF BIODEGRADABILITY STUDY NO. 6, UNIT NO. 4

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % $\frac{TVS}{TS}$ |
|---|---|---|---|---|---|---|---|
| 0 | 40.645 | 35.050 | 23.149 | - - | - - | - - | 65.9 |
| 2 | 38.725 | 33.015 | 21.278 | 4.72 | 5.81 | 8.08 | 64.5 |
| 5 | 26.616 | 31.730 | 18.438 | 34.52 | 9.47 | 20.35 | 58.1 |
| 12 | 19.725 | 24.089 | 13.199 | 51.47 | 31.27 | 42.98 | 54.8 |
| 19 | 18.107 | 22.644 | 11.715 | 55.45 | 35.4 | 49.39 | 51.7 |
| 27 | 15.422 | 20.938 | 10.046 | 62.06 | 40.26 | 56.60 | 48.0 |
| 31 | 1.366 | 1.563 | 1.204 | 96.64 | 95.54 | 94.80 | 77.0 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 52.61 | 43-47 |
| D.O., mg/ℓ | 4.42 | 0.2-5.8 |
| pH | 7.59 | 7.4-7.8 |
| Evaporation, mℓ/day | 400 | 200-600 |
| Airflow rate, ℓ/sec | .0657 | .047-.079 |

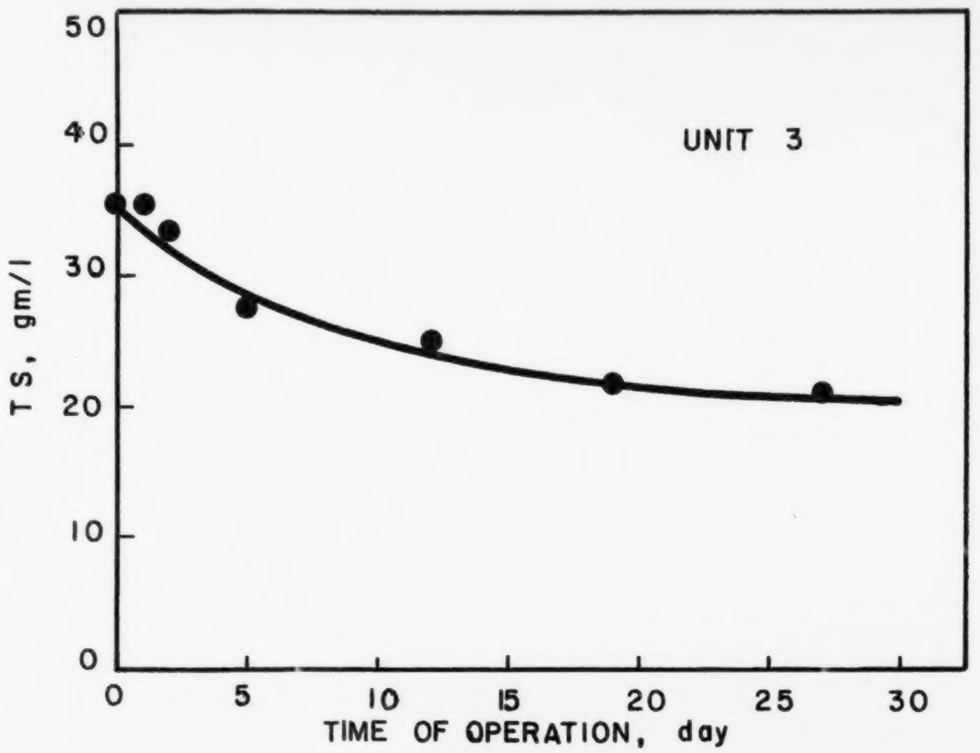

Figure C-69. Total solids (TS) removal as related to time of operation, biodegradability study no. 6.

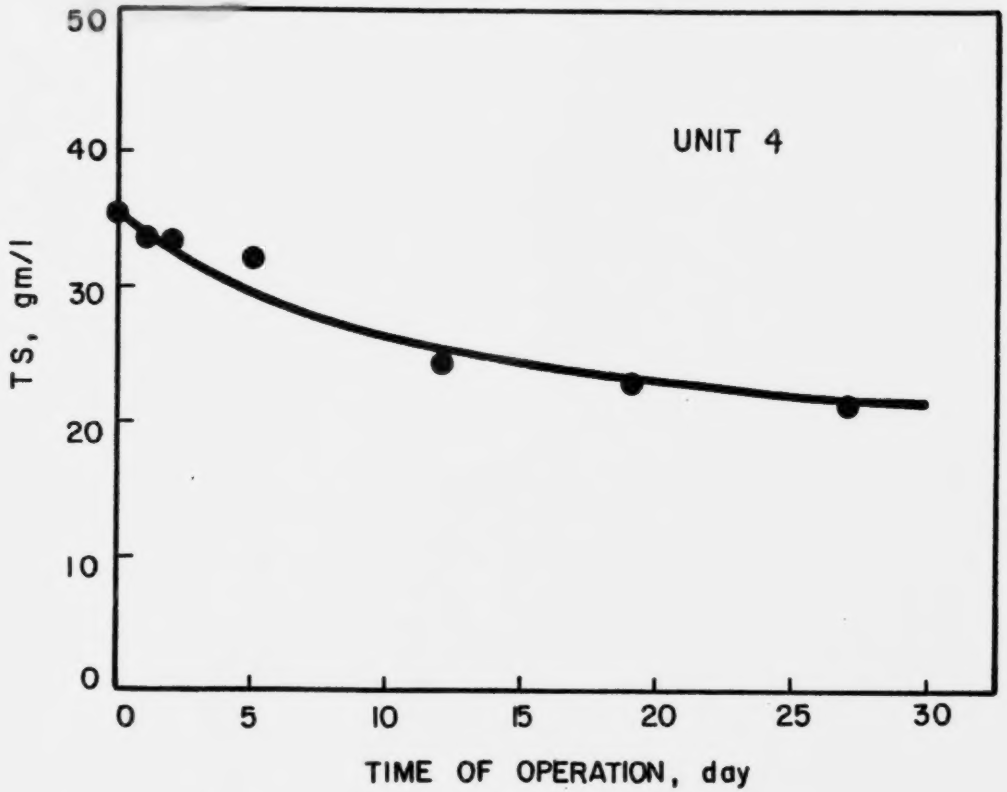

Figure C-70. Total solids (TS) removal as related to time of operation, biodegradability study no. 6.

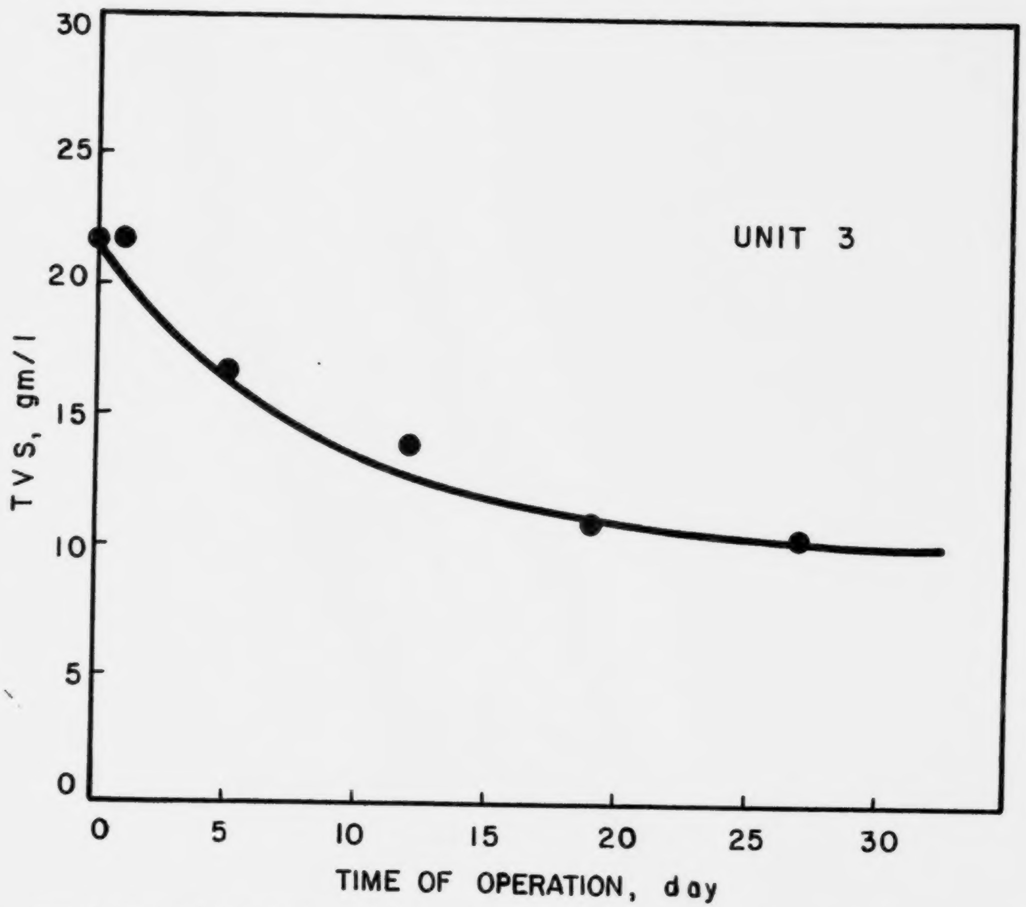

Figure C-71. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 6.

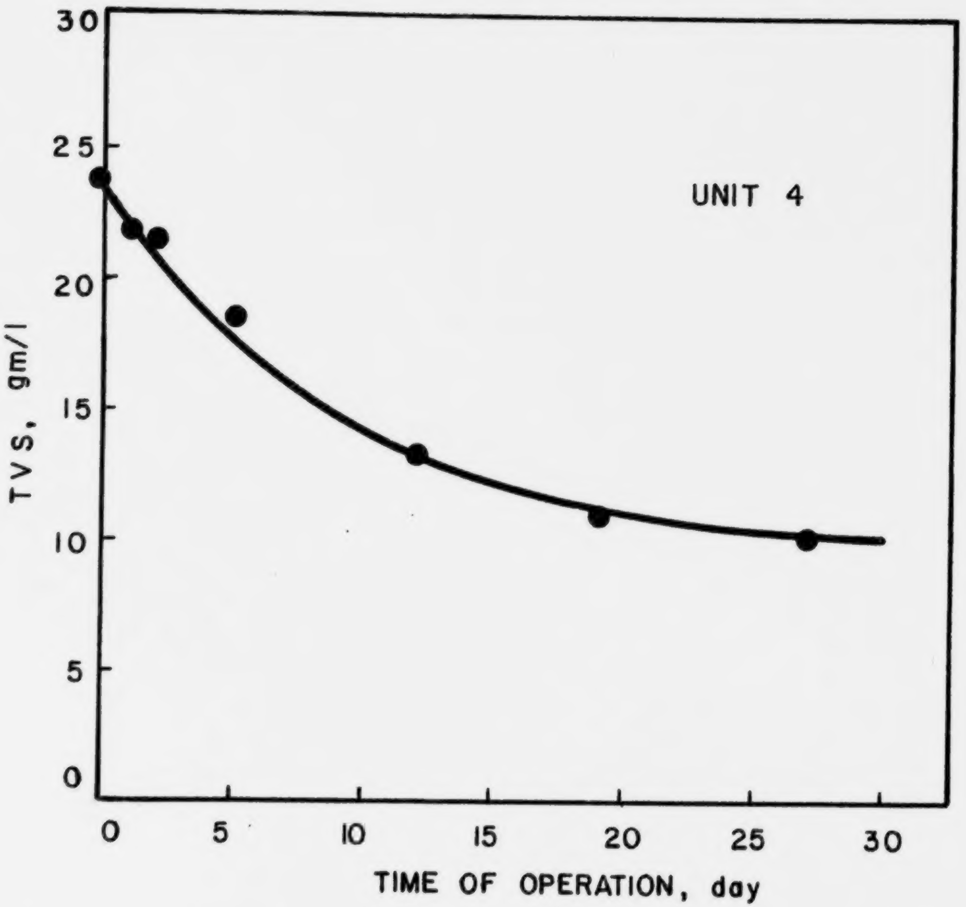

Figure C-72. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 6.

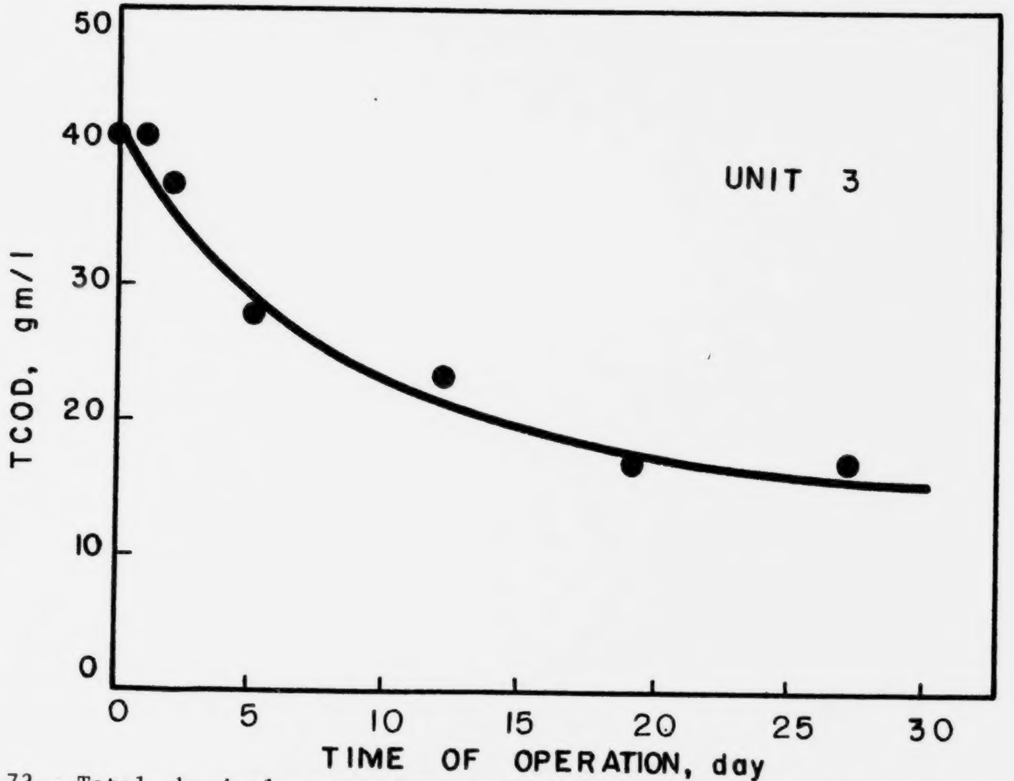

Figure C-73. Total chemical oxygen demand (TCOD) removal as related to time of operation. biodegradability study no. 6.

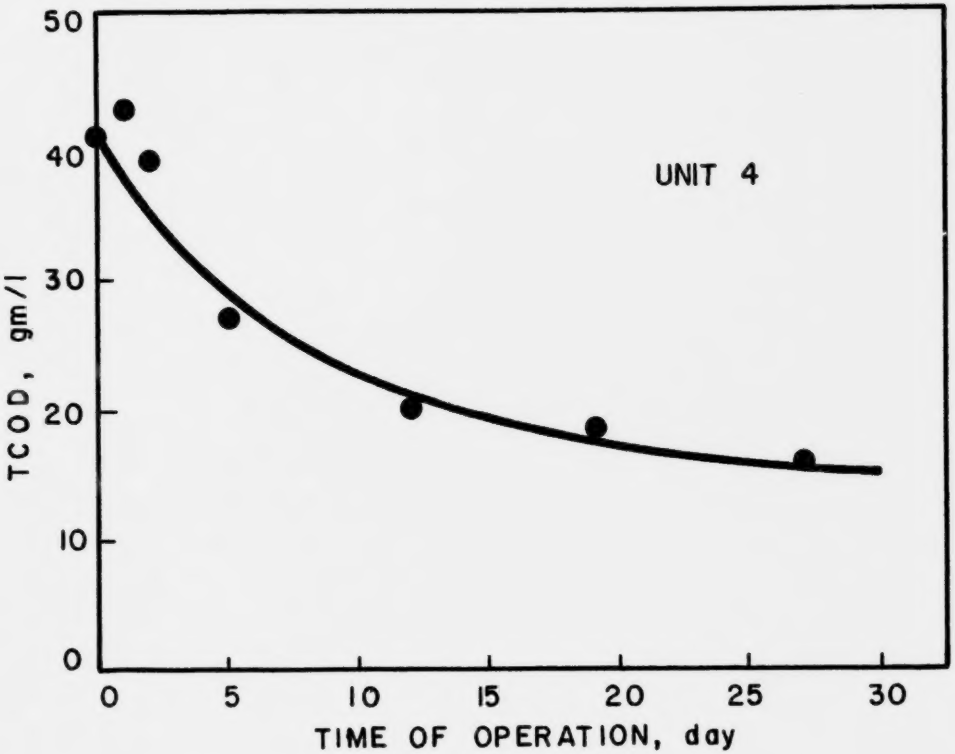

Figure C-74. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 6.

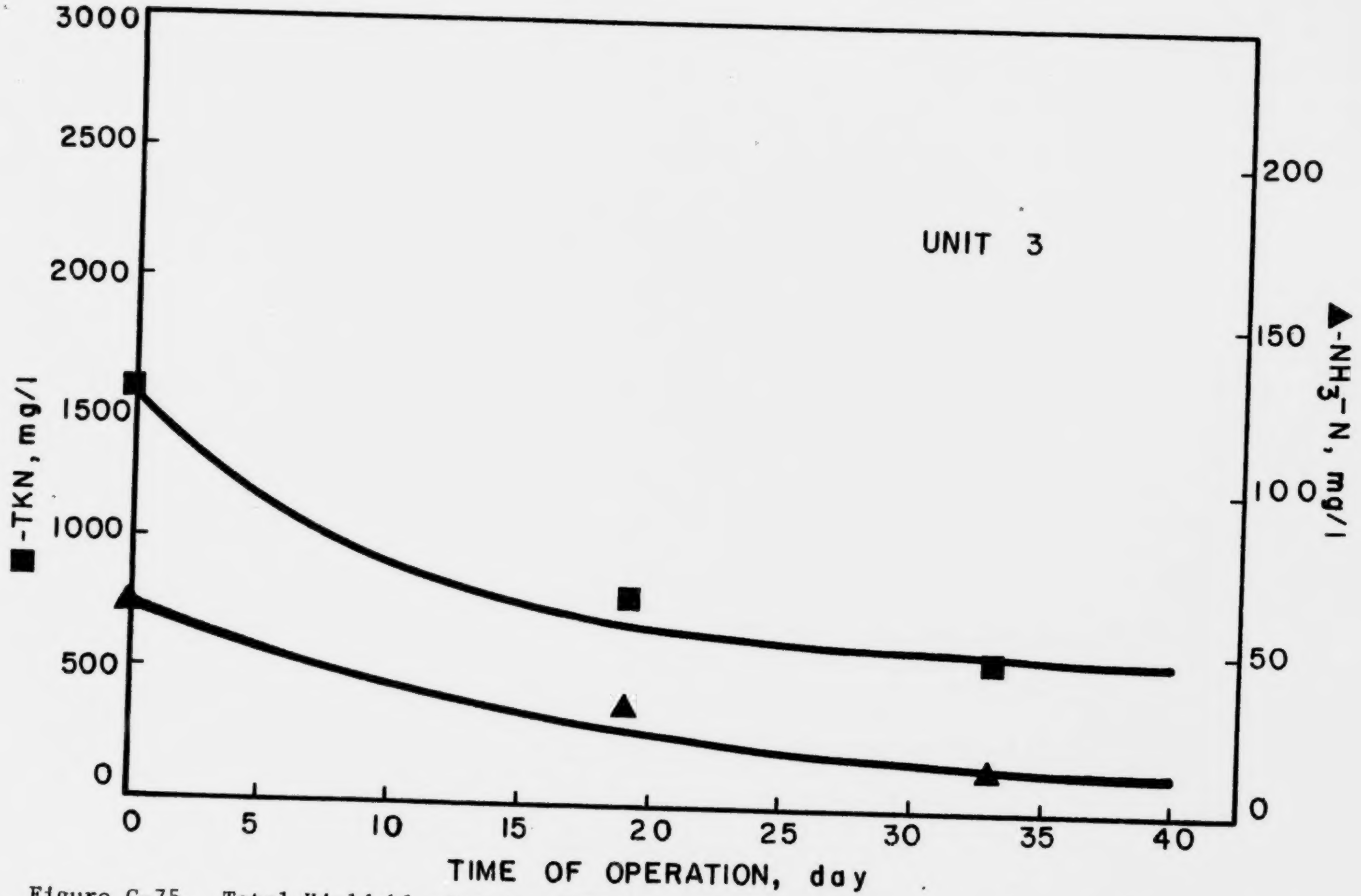

Figure C-75. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 6.

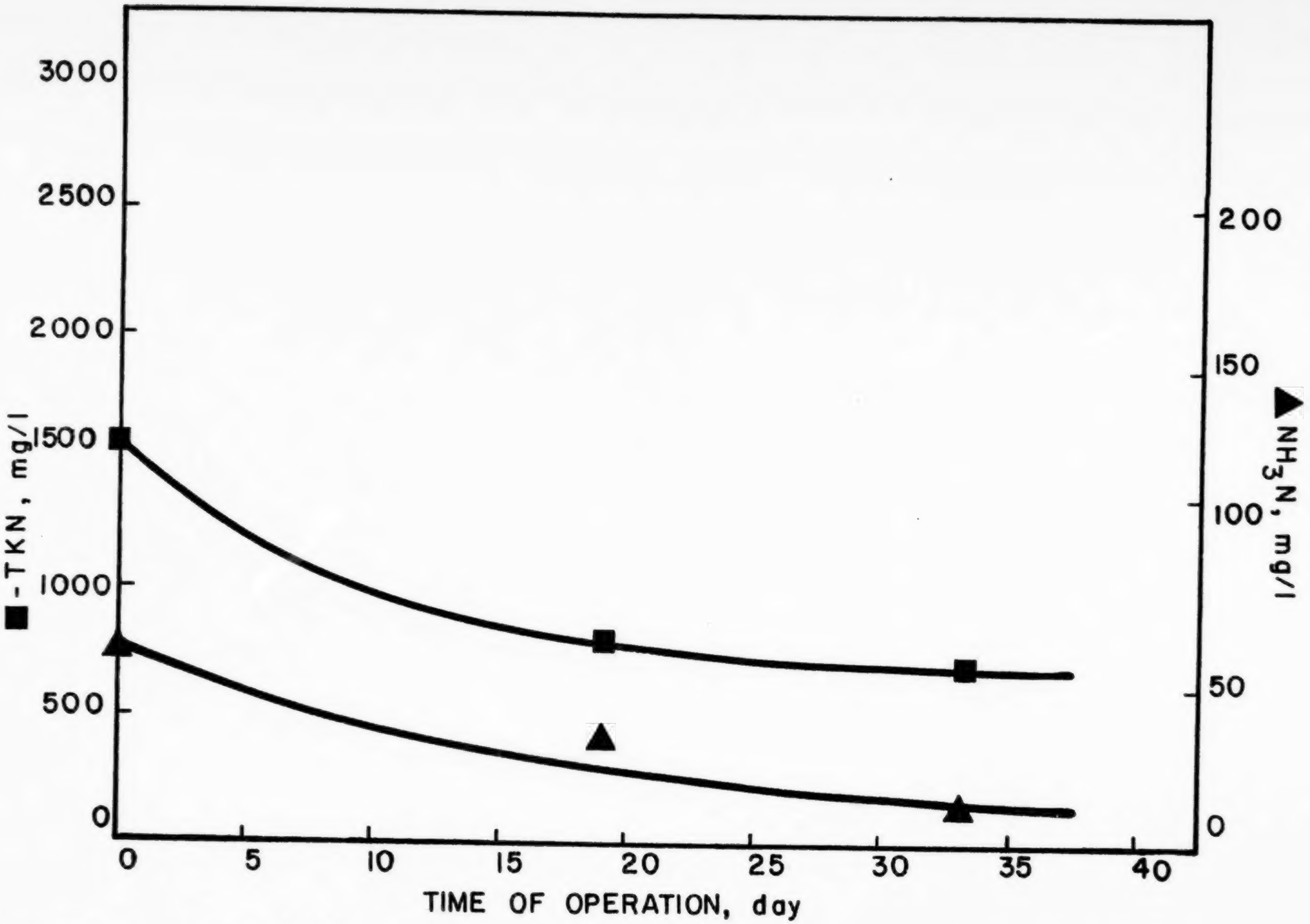

Figure C-76. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 6.

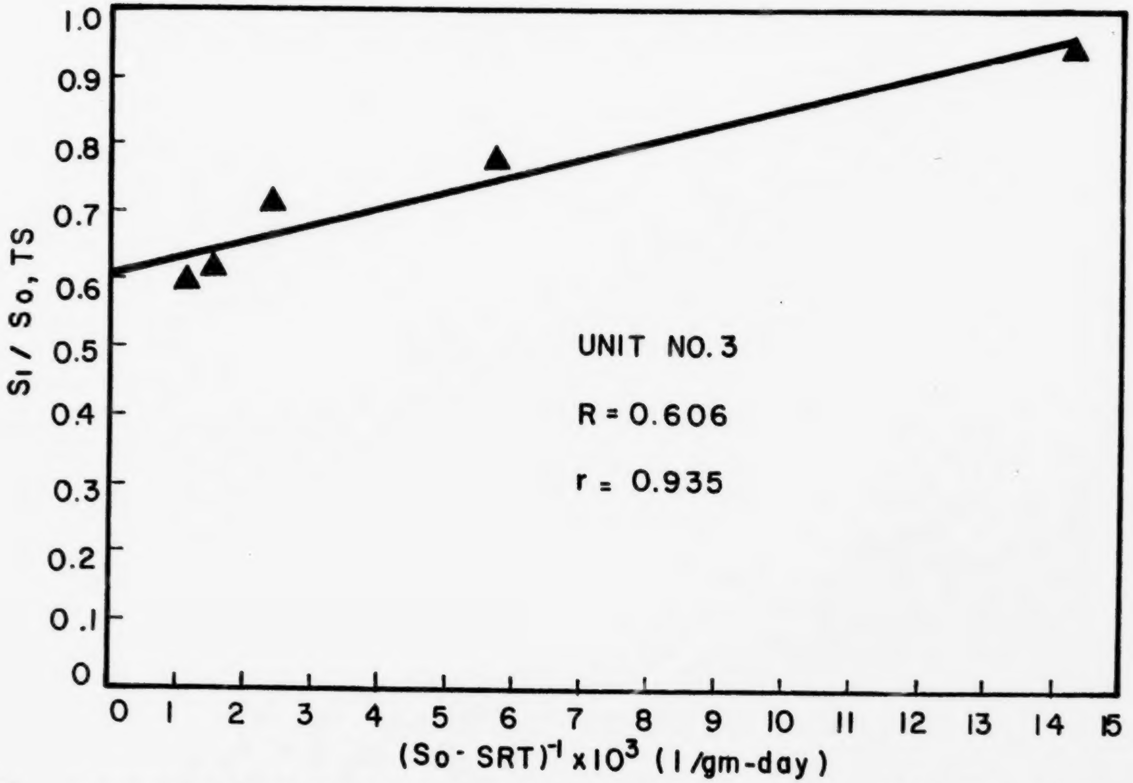

Figure C-77. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 6.

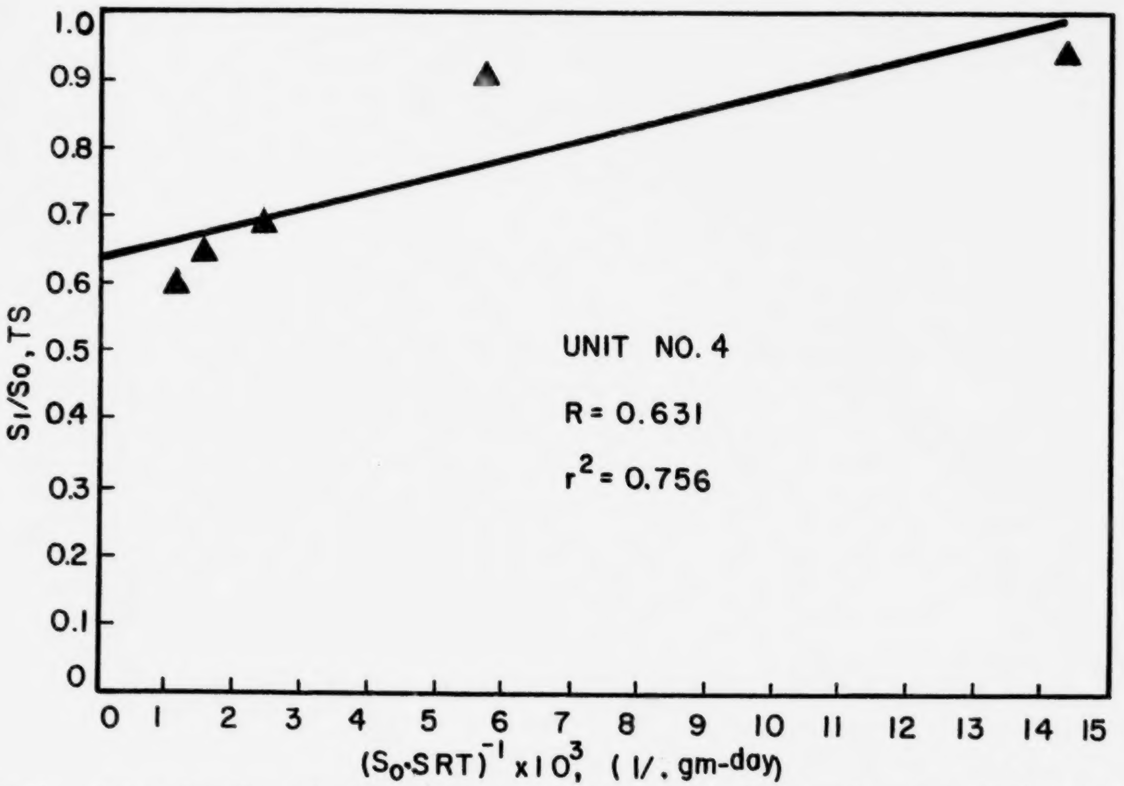

Figure C-78. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 6.

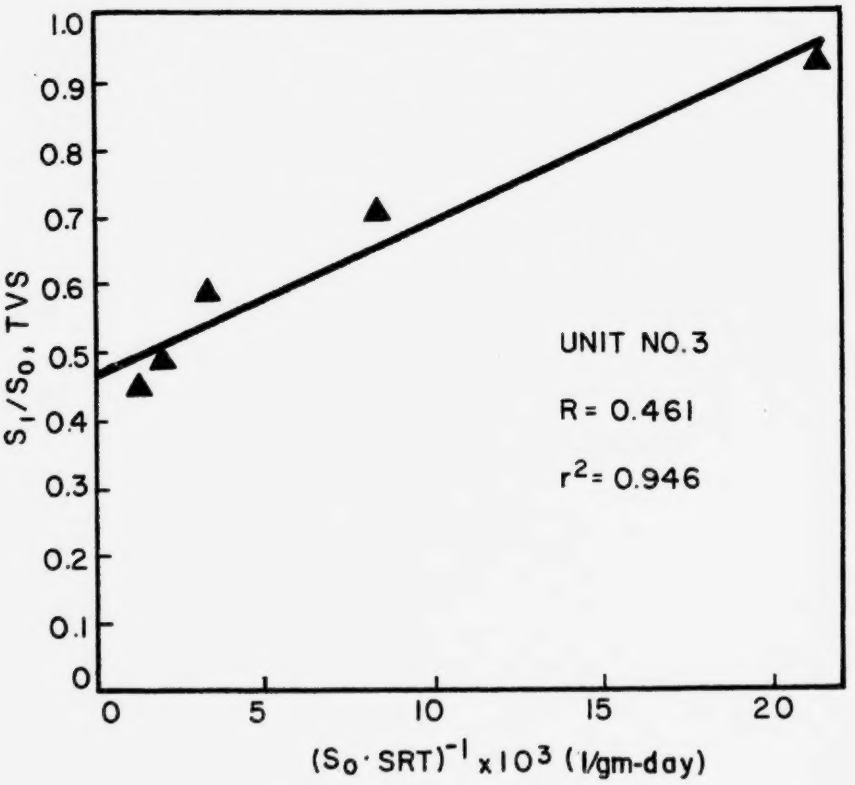

Figure C-79. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 6.

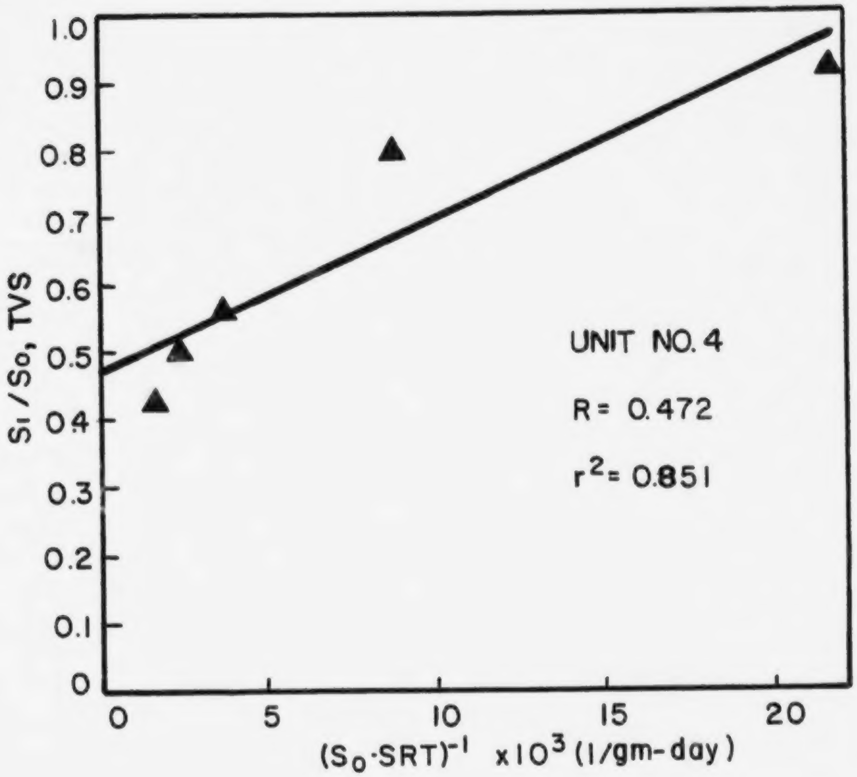

Figure C-80. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 6.

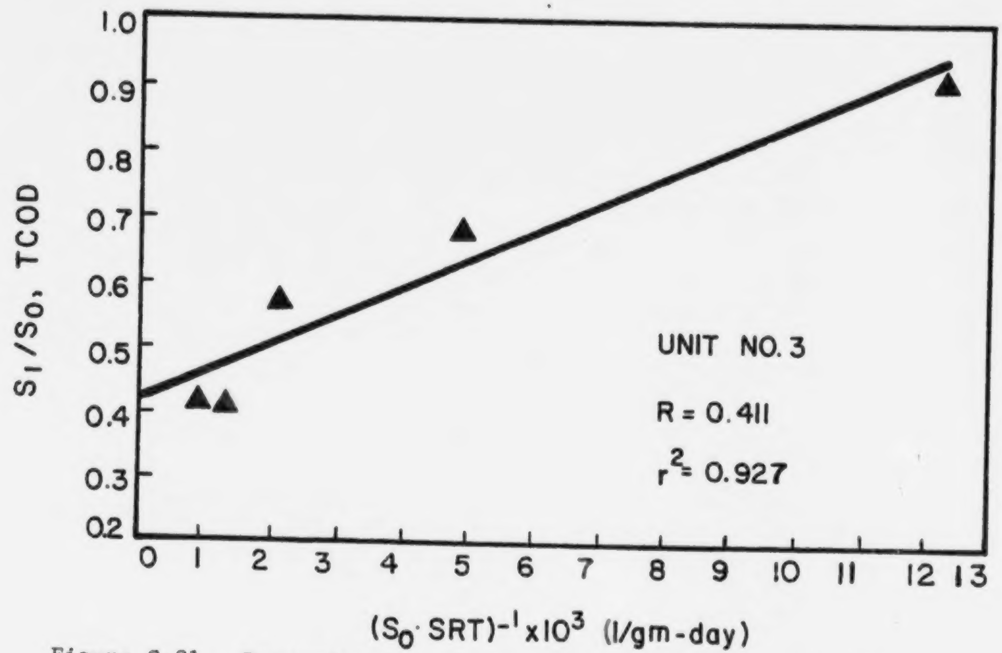

Figure C-81. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 6.

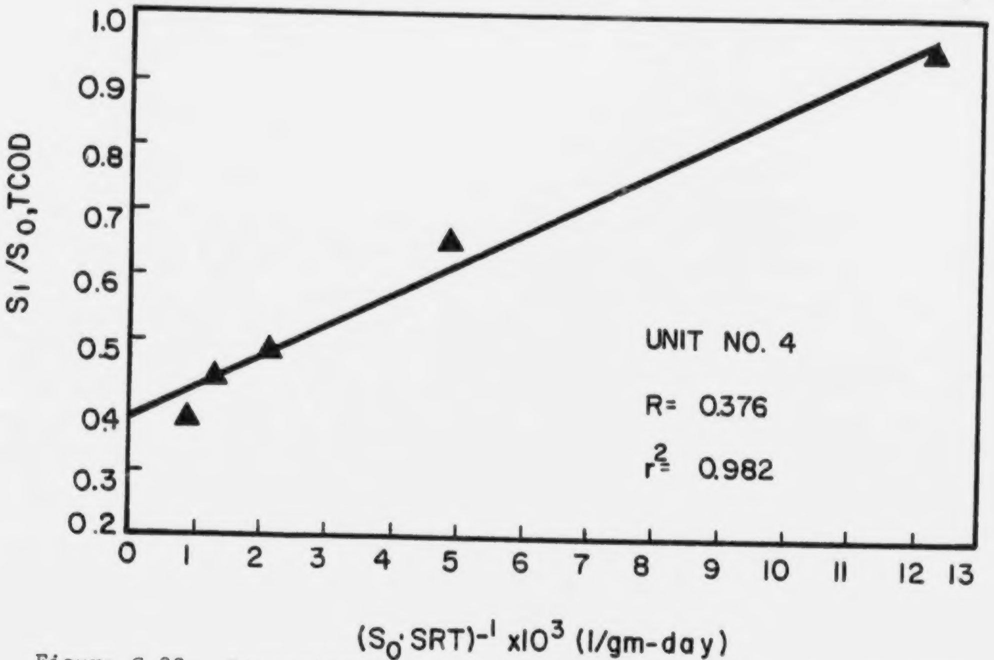

Figure C-82. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 6.

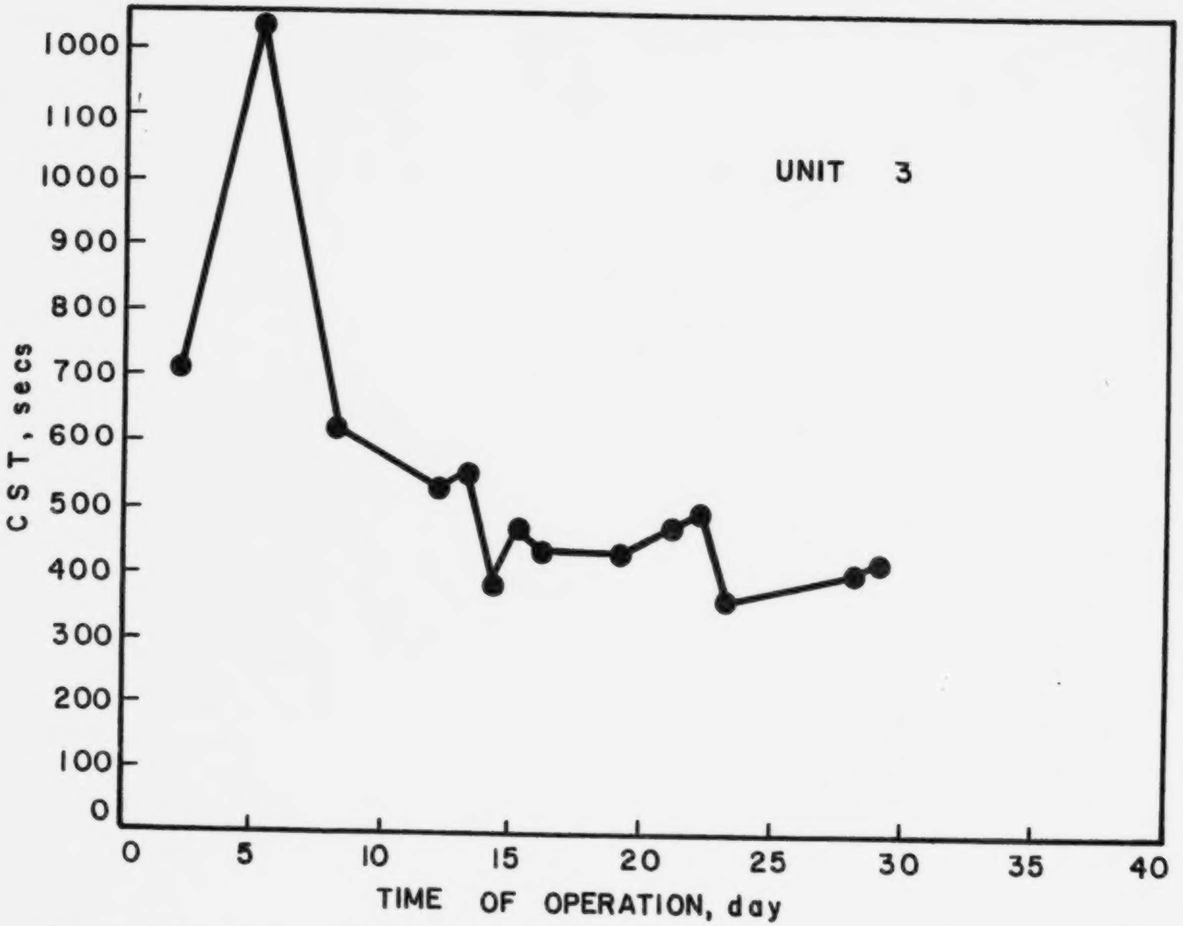

Figure C-83. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 6.

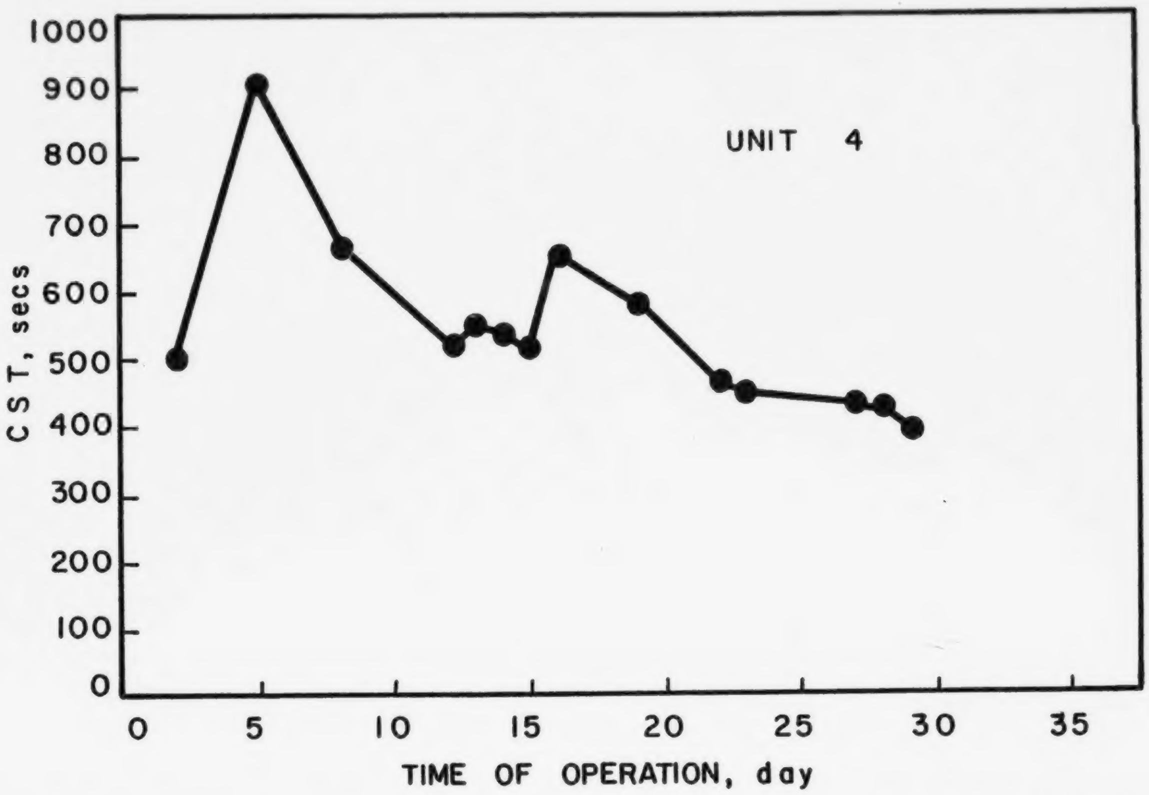

Figure C-84. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 6.

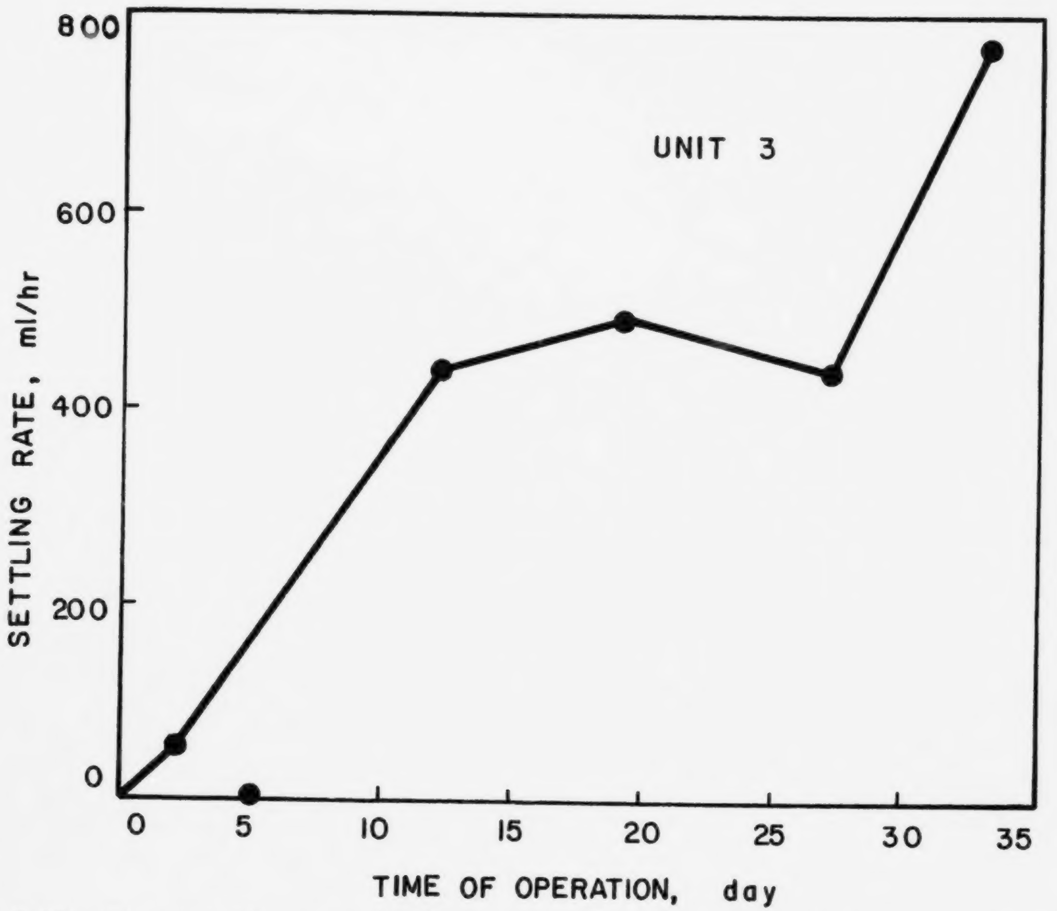

Figure C-85. Zone settling rate (ZSR) as related to time of operation, biodegradability study no. 6.

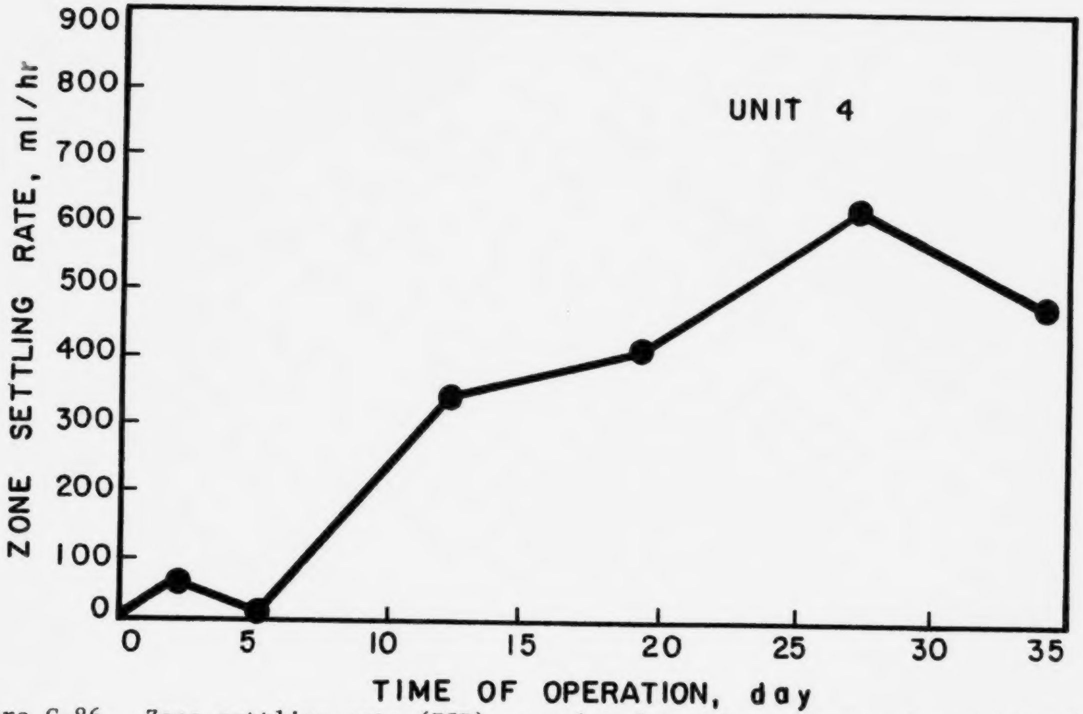

Figure C-86. Zone settling rate (ZSR) as related to time of operation, biodegradability study no. 6.

TABLE C-13. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 7, UNIT NO. 1

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % $\frac{TVS}{TS}$ |
|---|---|---|---|---|---|---|---|
| 0 | 37.949 | 35.074 | 22.242 | - - | - - | - - | 63.4 |
| 7 | 30.367 | 37.160 | 20.548 | 19.9 | - - | 7.61 | 53.3 |
| 21 | 17.585 | 29.930 | 14.184 | 53.7 | 14.7 | 36.2 | 47.4 |
| 27 | 16.188 | 28.105 | 12.231 | 57.3 | 19.9 | 45.0 | 43.6 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 50.1 | 47-54 |
| D.O., mg/ℓ | 3.24 | 0.15-4.7 |
| pH | 7.2 | 5.5-7.8 |
| Evaporation, mℓ/day | 450 | 400-500 |
| Airflow rate, ℓ/sec | 0.07 | - - |

TABLE C-14. RESULTS OF BATCH BIODEGRADABILITY STUDY NO. 7, UNIT NO. 2

| Day of Operation | TCOD gm/ℓ | TS gm/ℓ | TVS gm/ℓ | % Removal TCOD | % Removal TS | % Removal TVS | % TVS/TS |
|---|---|---|---|---|---|---|---|
| 0 | 37.949 | 35.074 | 22.242 | - - | - - | - - | 63.4 |
| 2 | 36.991 | 32.743 | 20.202 | 2.52 | 6.64 | 25.8 | 61.7 |
| 7 | 23.783 | 29.439 | 16.078 | 37.3 | 16.1 | 40.9 | 54.6 |
| 21 | 15.766 | 24.277 | 11.379 | 58.4 | 30.8 | 58.2 | 46.9 |
| 27 | 17.233 | 22.551 | 10.020 | 54.6 | 35.7 | 63.2 | 44.4 |

| Parameter | Average | Range |
|---|---|---|
| Reactor Temperature, °C | 50.275 | 47-54 |
| D.O., mg/ℓ | 3.434 | .13-4.8 |
| pH | 7.375 | 5.5-7.9 |
| Evaporation, mℓ/day | 450 | 400-500 |
| Airflow rate, ℓ/sec | 0.071156 | 0.063-0.079 |

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CENTER OF PAGE
TOP OF IMAGE AREA
DROPPED HEAD; BEGIN SECTIONS HERE
1/2"
9-1/8"
BEGIN LAST LINE OF TEXT
BOTTOM OF IMAGE AREA OUTSIDE DIMENSION FOR TABLES AND ILLUSTRATIONS
3/8"
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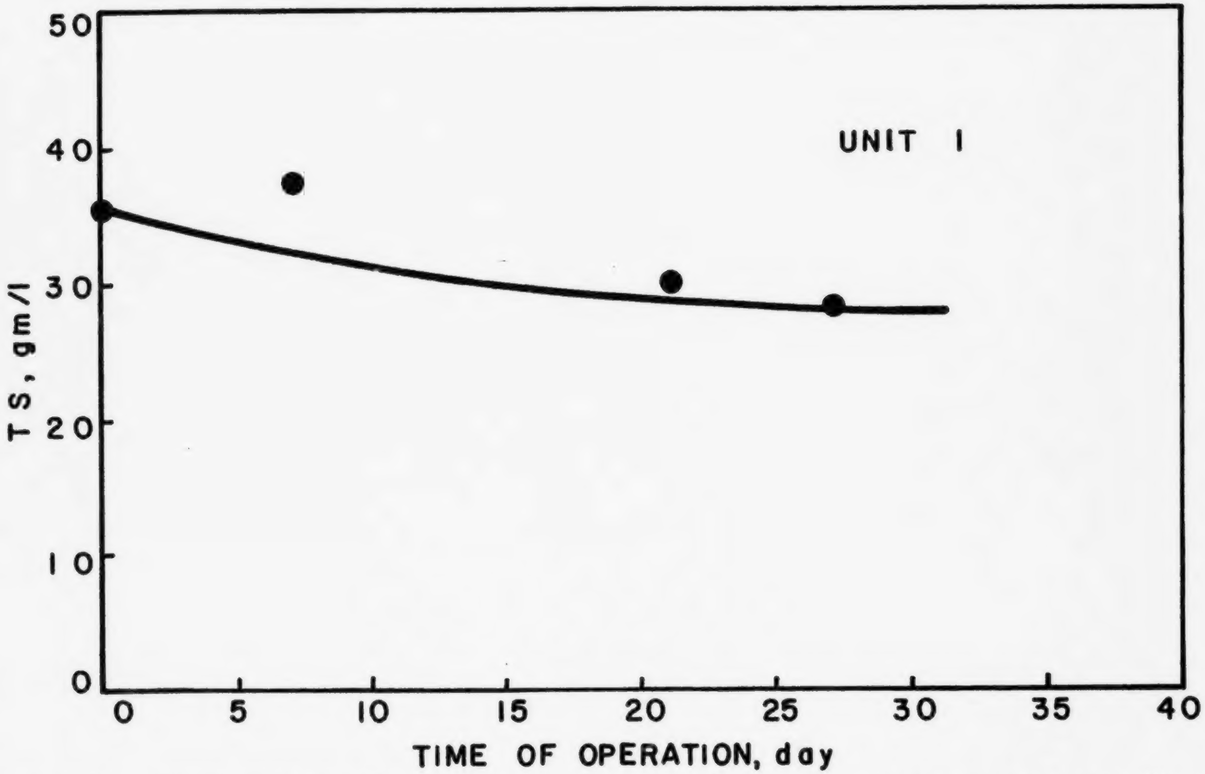

Figure C-87. Total solids (TS) removal as related to time of operation, biodegradability study no. 7.

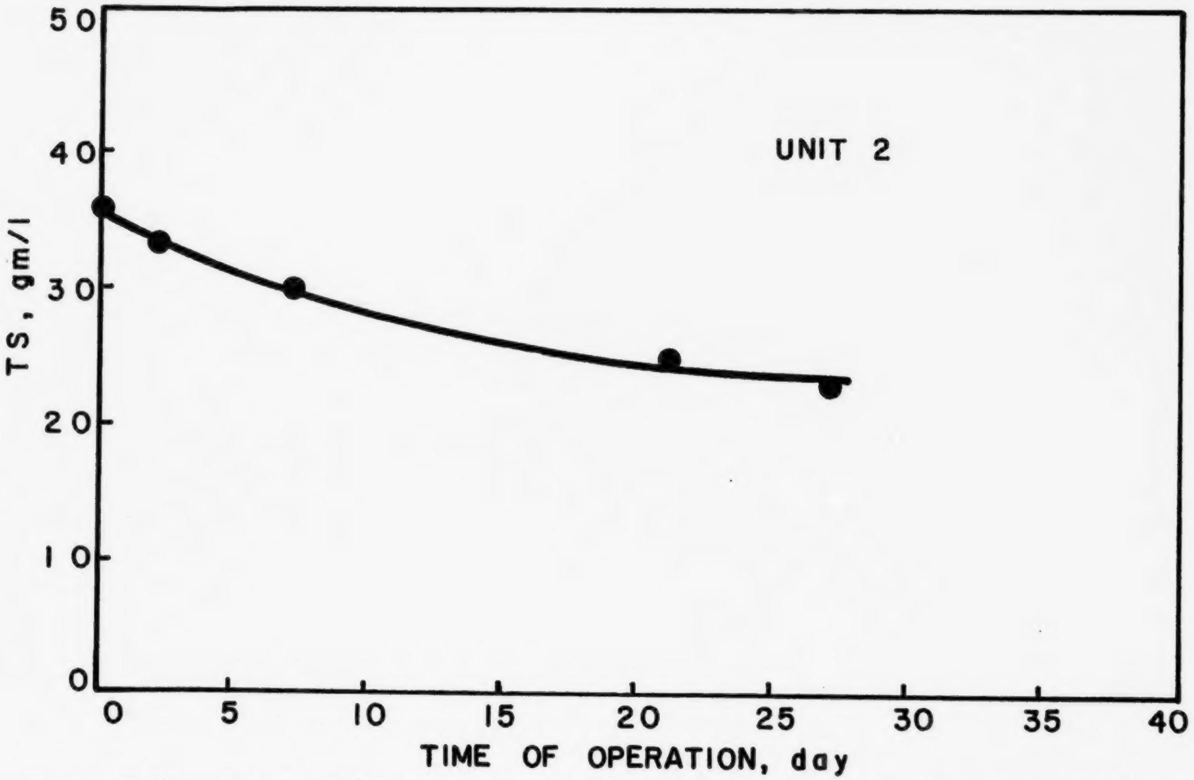

Figure C-88. Total solids (TS) removal as related to time of operation, biodegradability study no. 7.

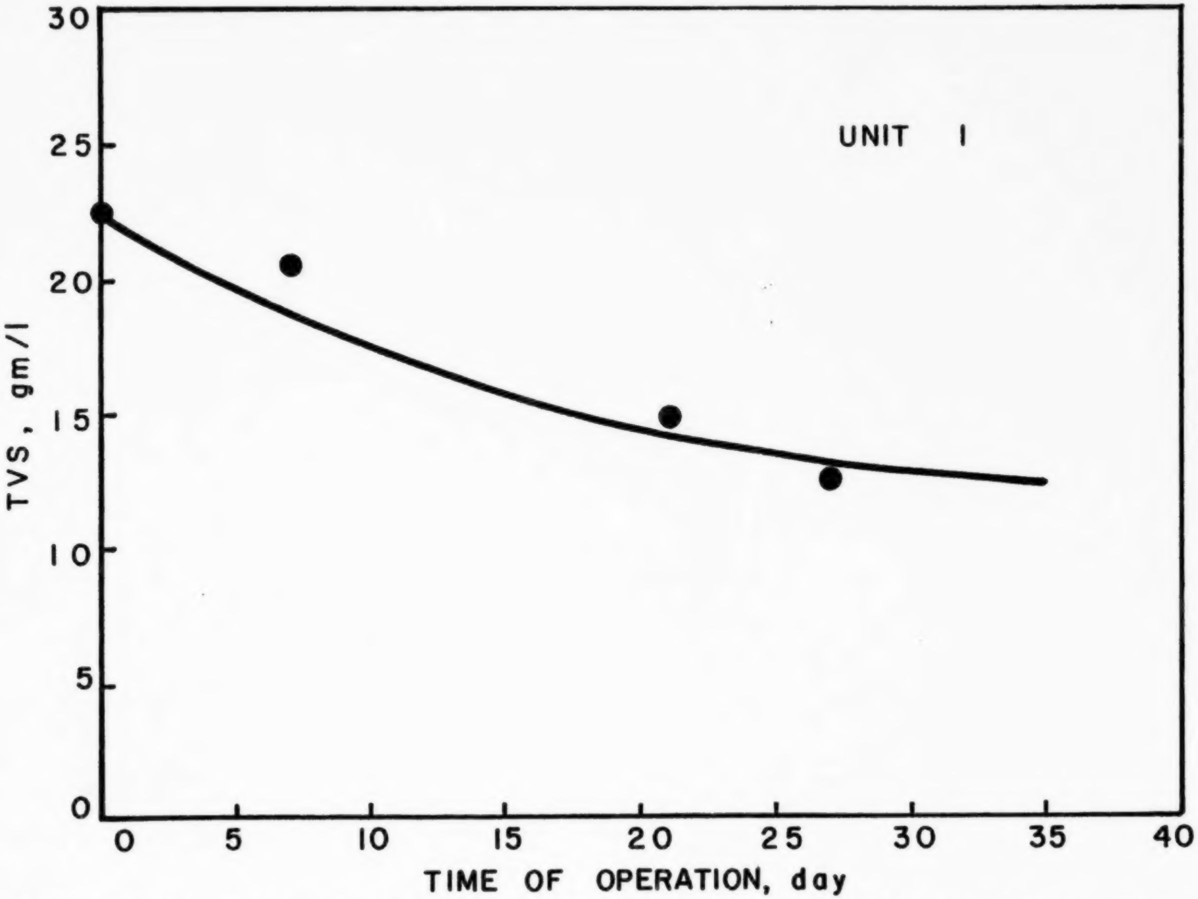

Figure C-89. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 7.

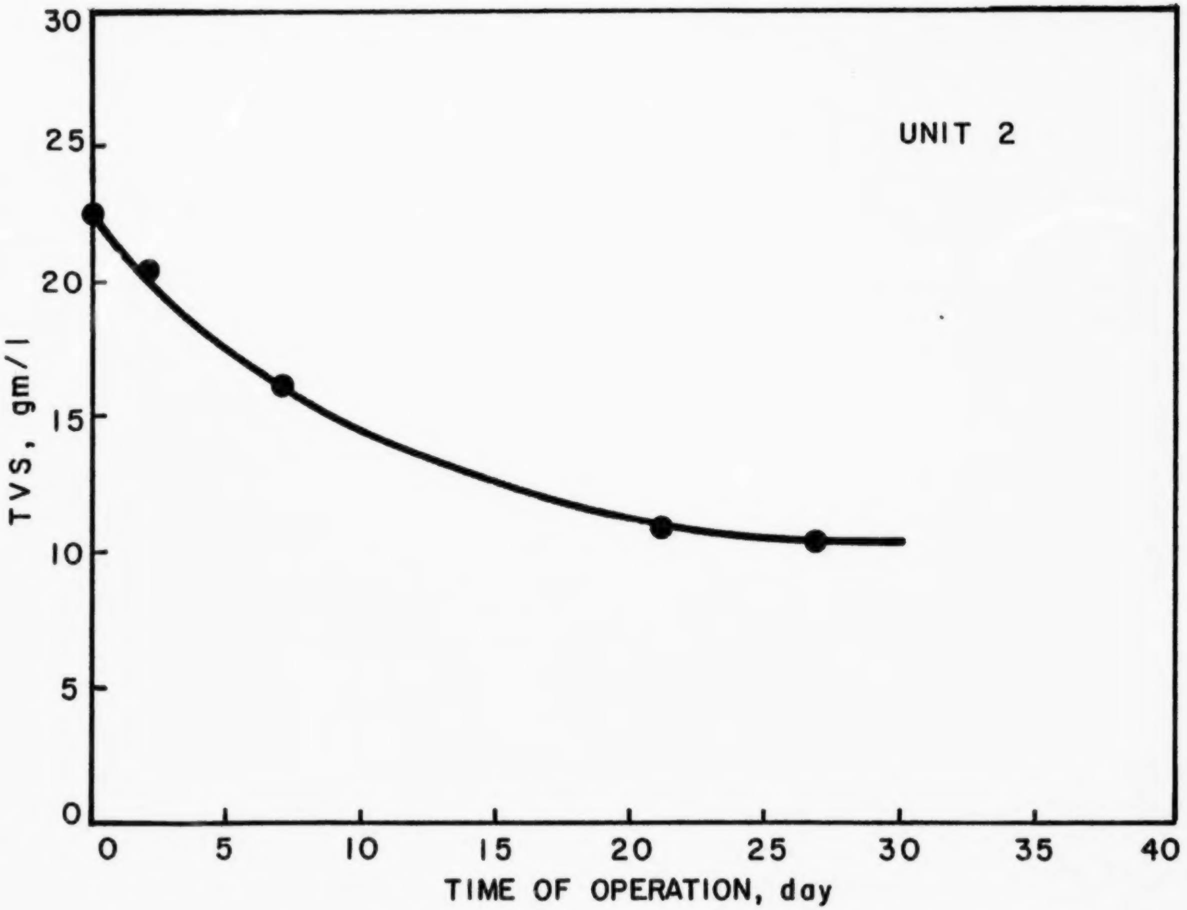

Figure C-90. Total volatile solids (TVS) removal as related to time of operation, biodegradability study no. 7.

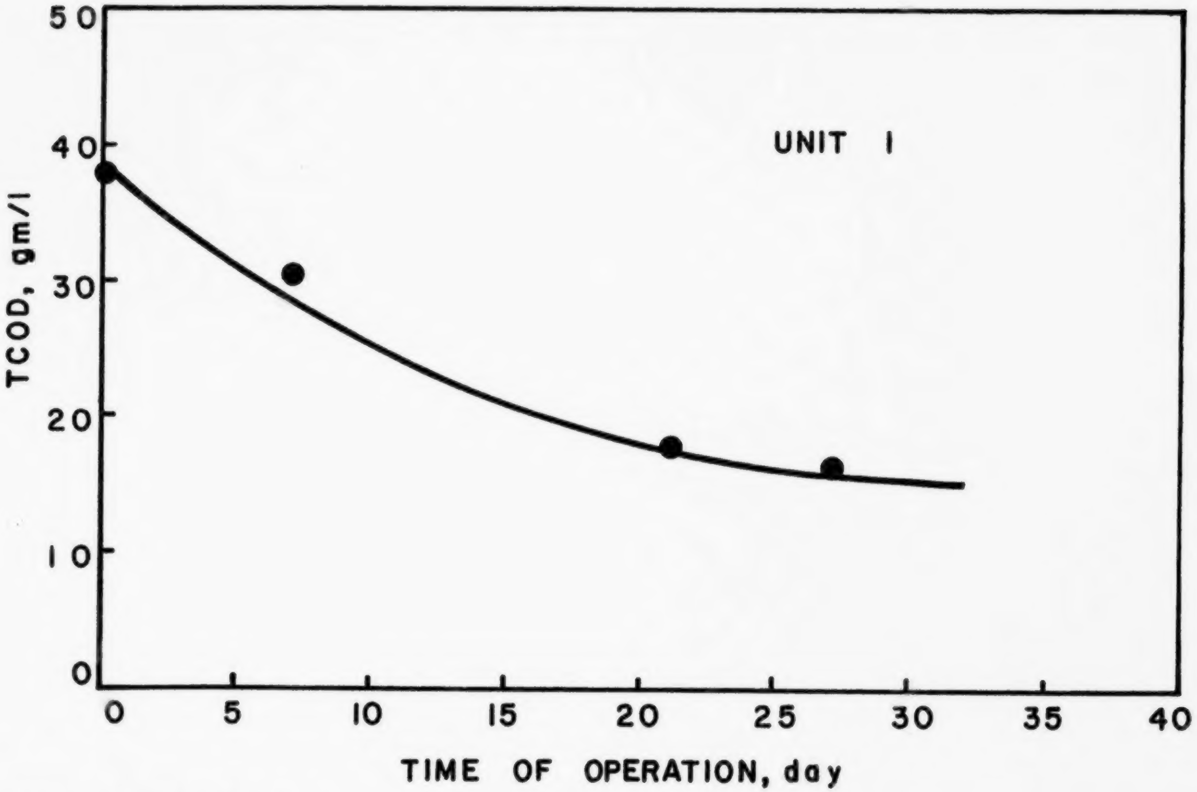

Figure C-91. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 7.

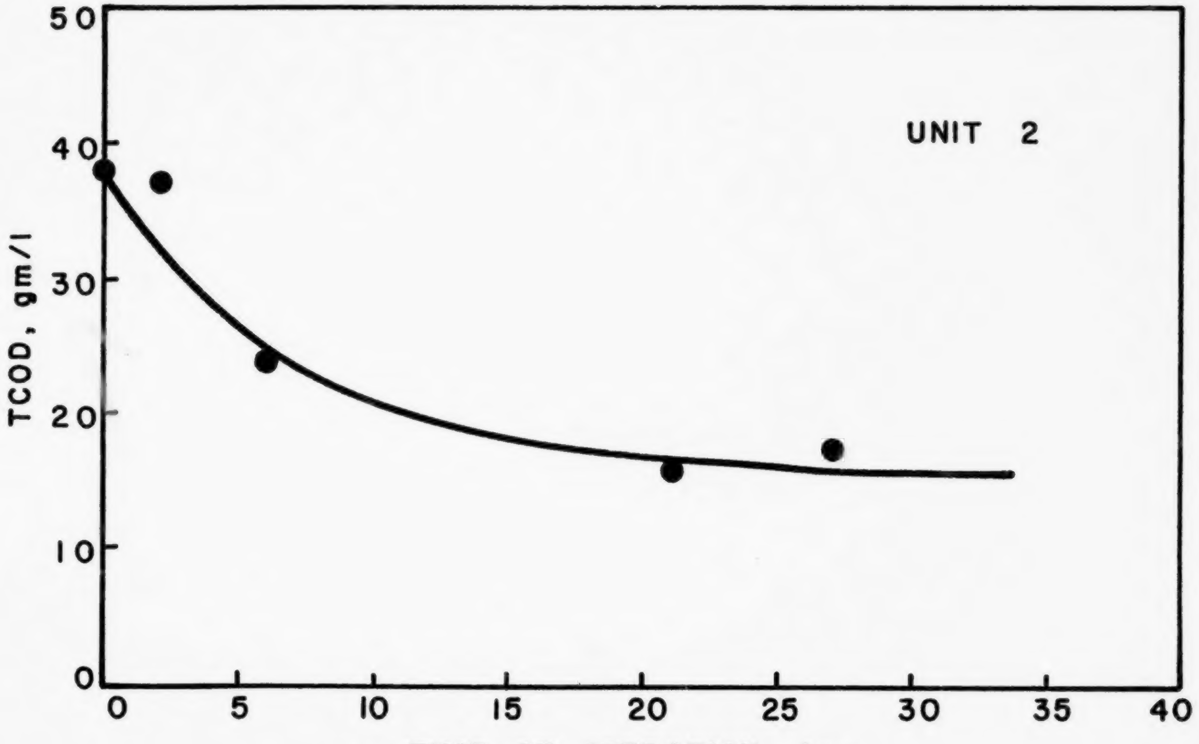

Figure C-92. Total chemical oxygen demand (TCOD) removal as related to time of operation, biodegradability study no. 7.

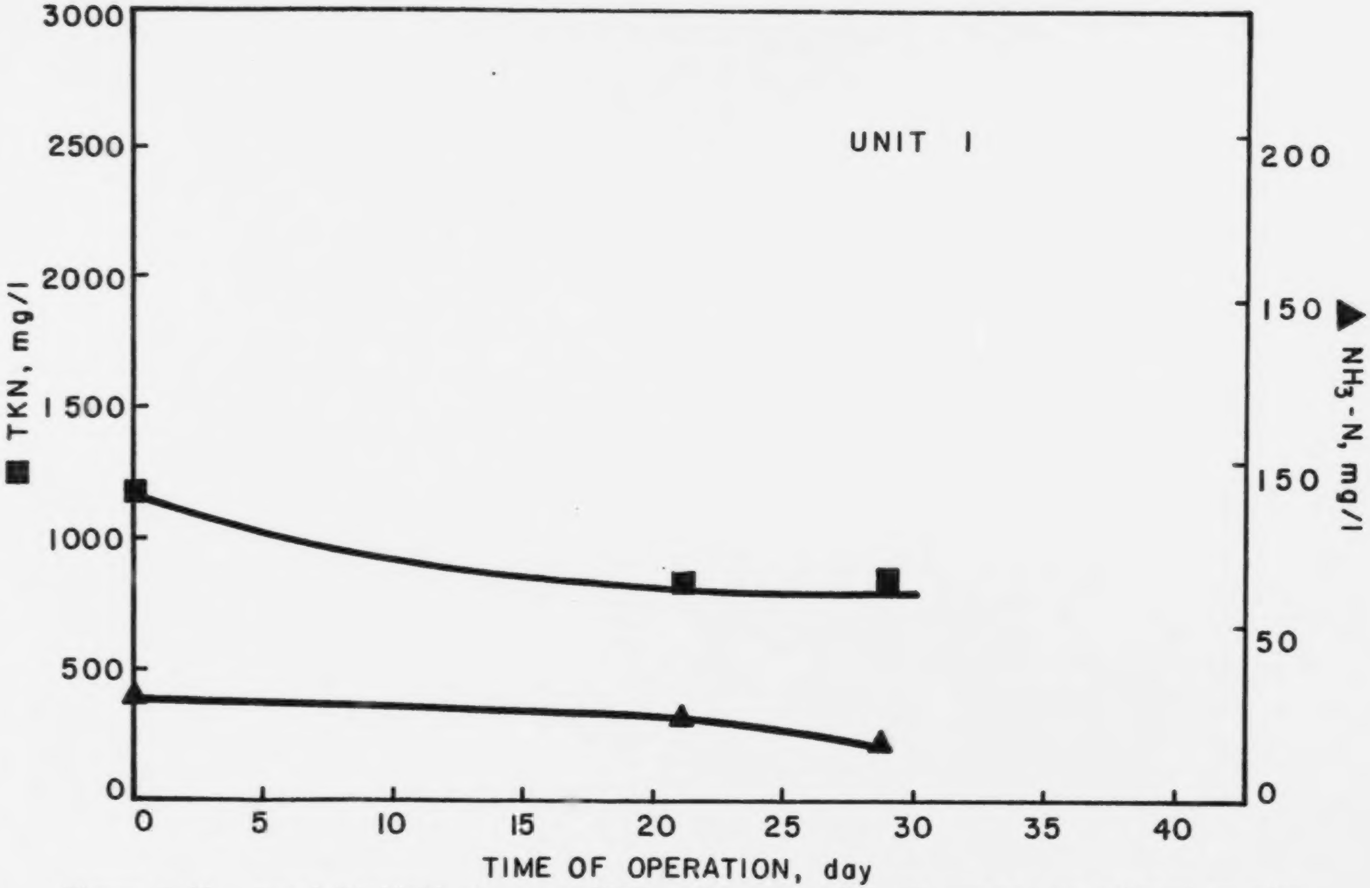

Figure C-93. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 7.

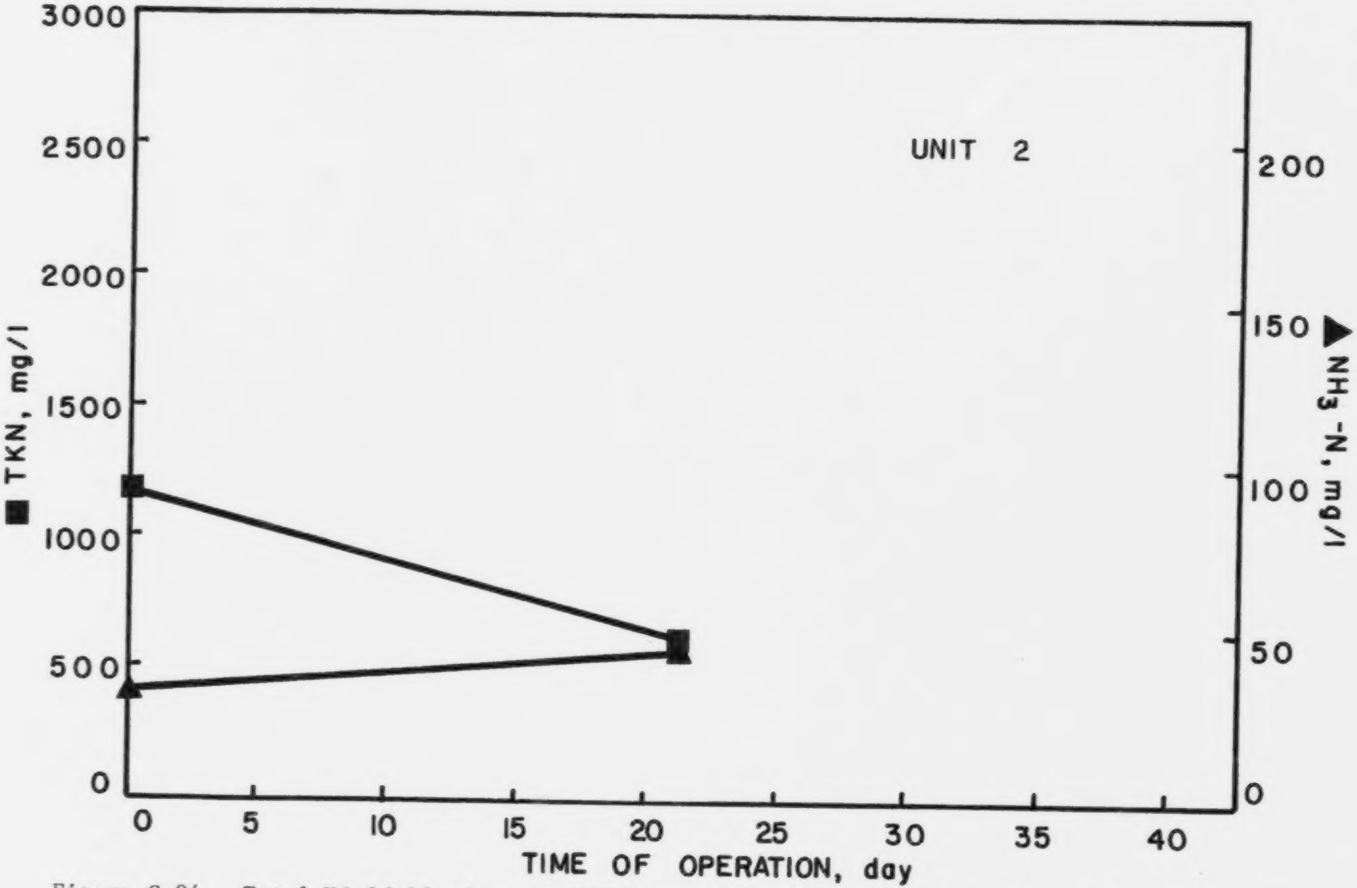

Figure C-94. Total Kjeldahl nitrogen (TKN) and ammonia ($NH_3$-N) removal as related to time of operation, biodegradability study no. 7.

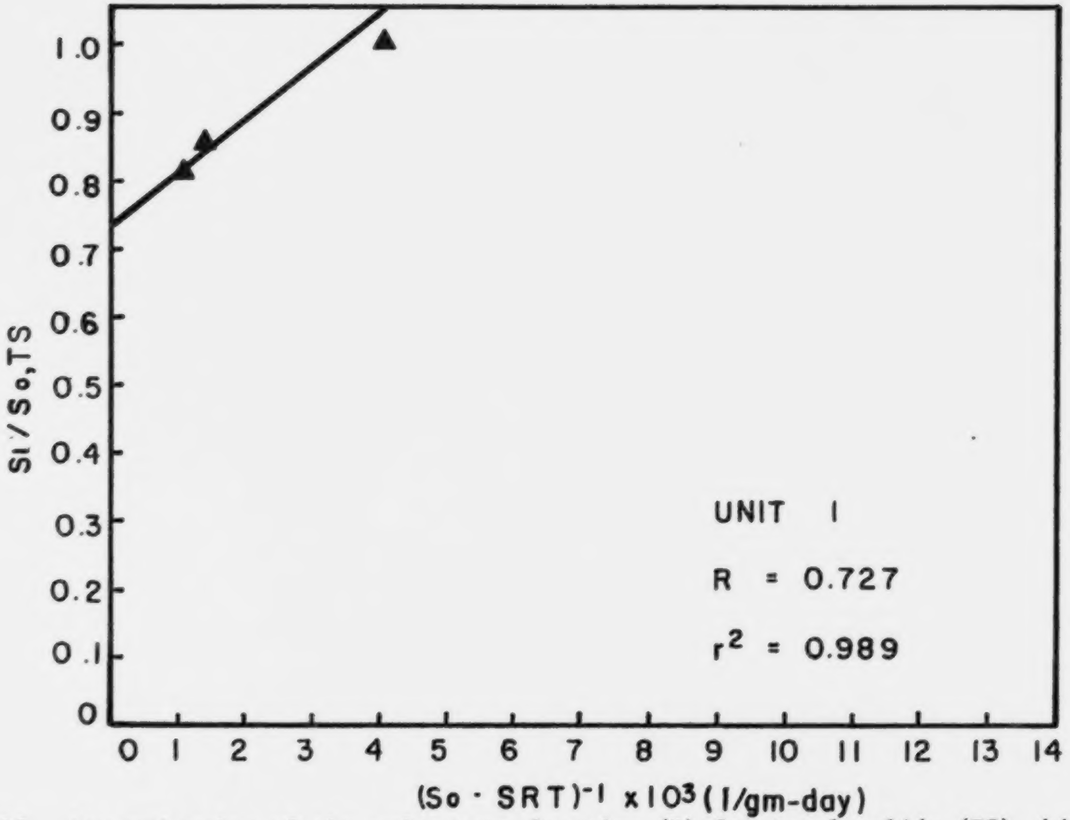

Figure C-95. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 7.

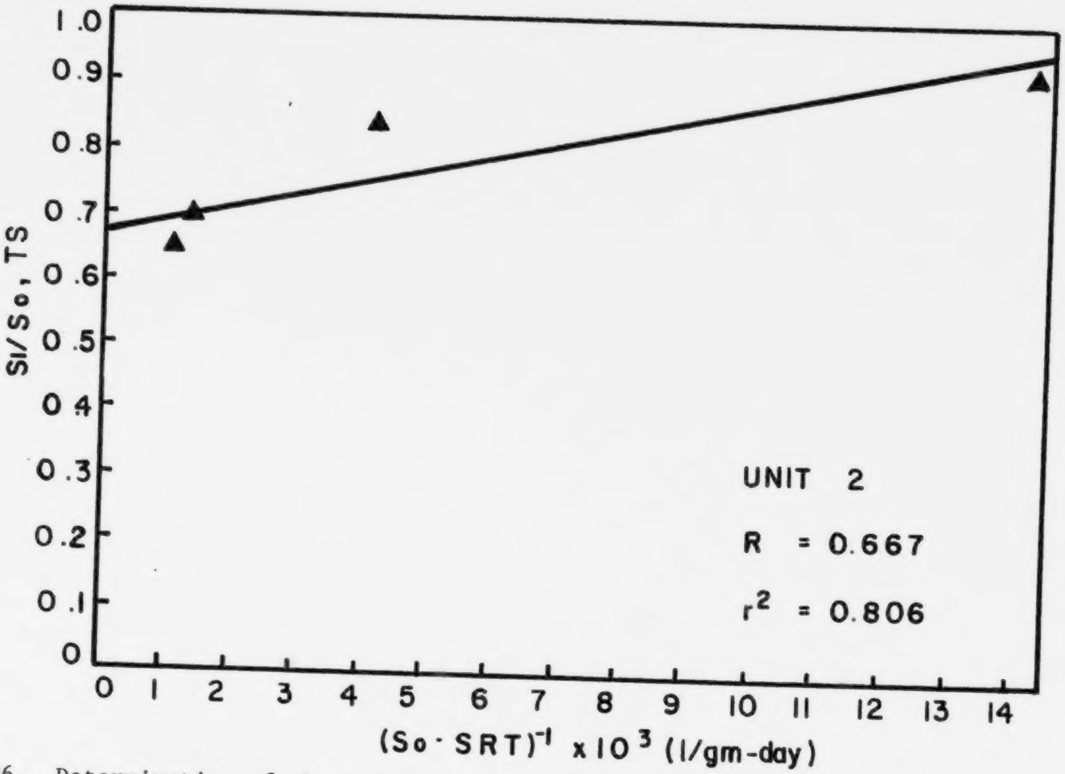

Figure C-96. Determination of the refractory fraction (R) for total solids (TS), biodegradability study no. 7.

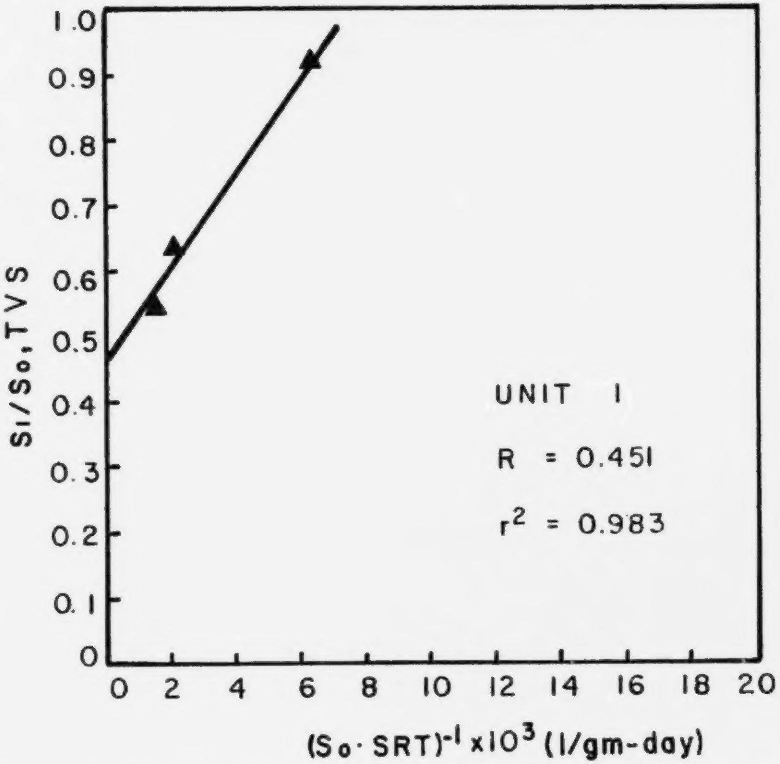

Figure C-97. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 7.

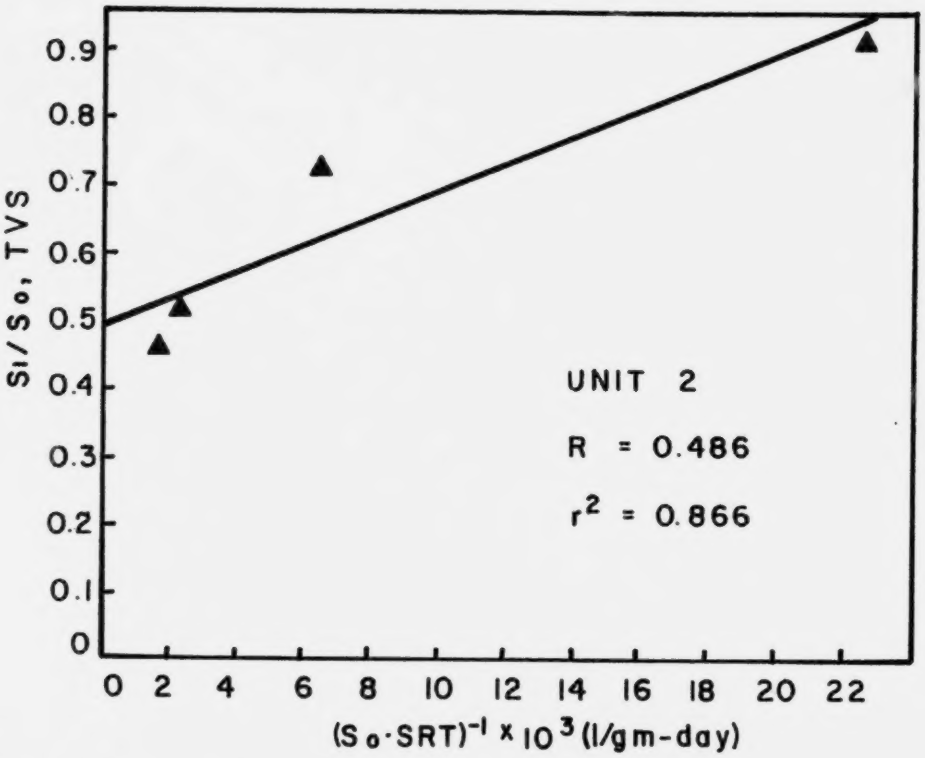

Figure C-98. Determination of the refractory fraction (R) for total volatile solids (TVS), biodegradability study no. 7.

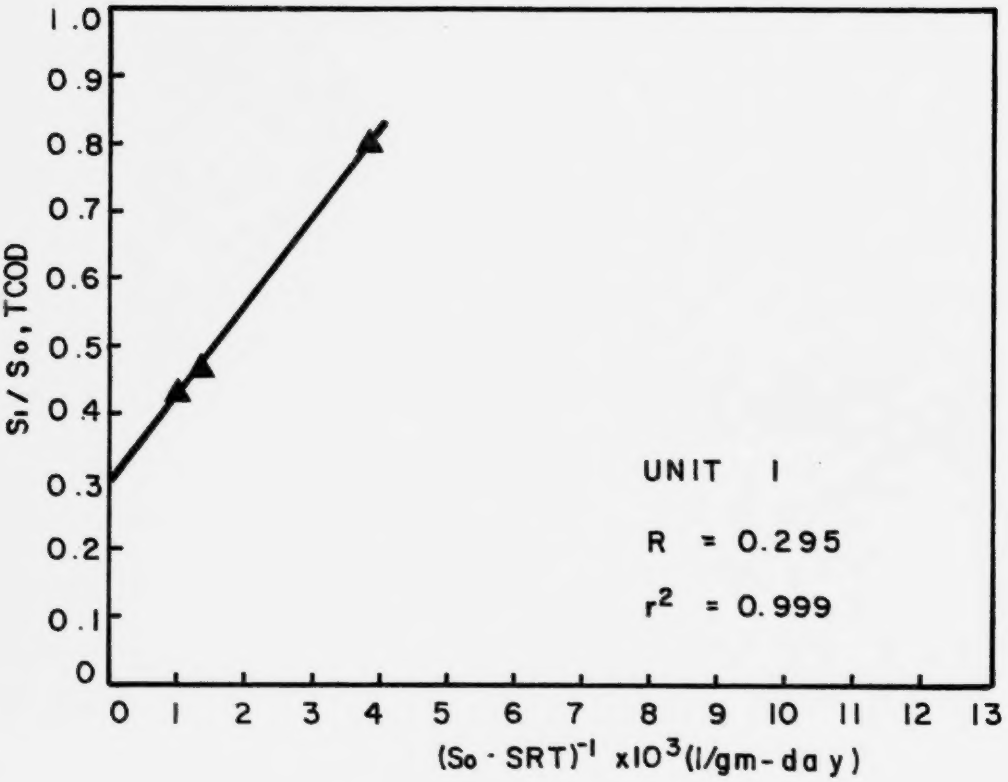

Figure C-99. Determination of the refractory fraction (R) for total chemical oxygen demand (TCOD), biodegradability study no. 7.

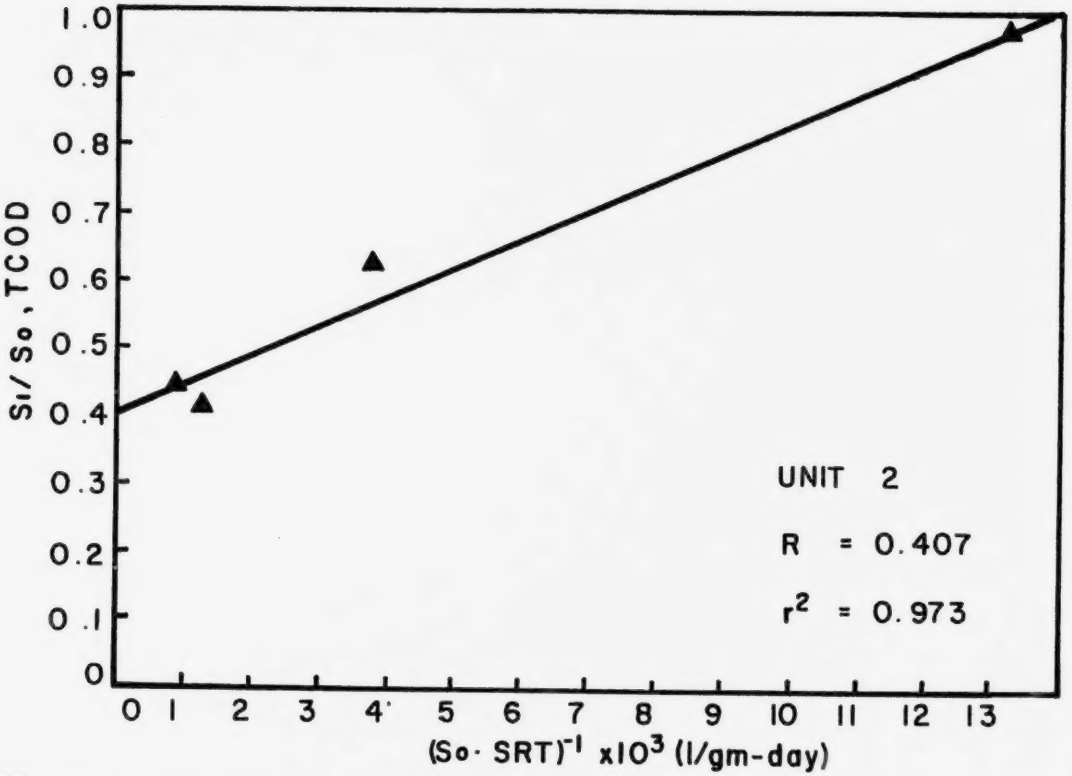

Figure C-100. Determination of the refractory fraction (R), for total chemical oxygen demand (TCOD), biodegradability study no. 7.

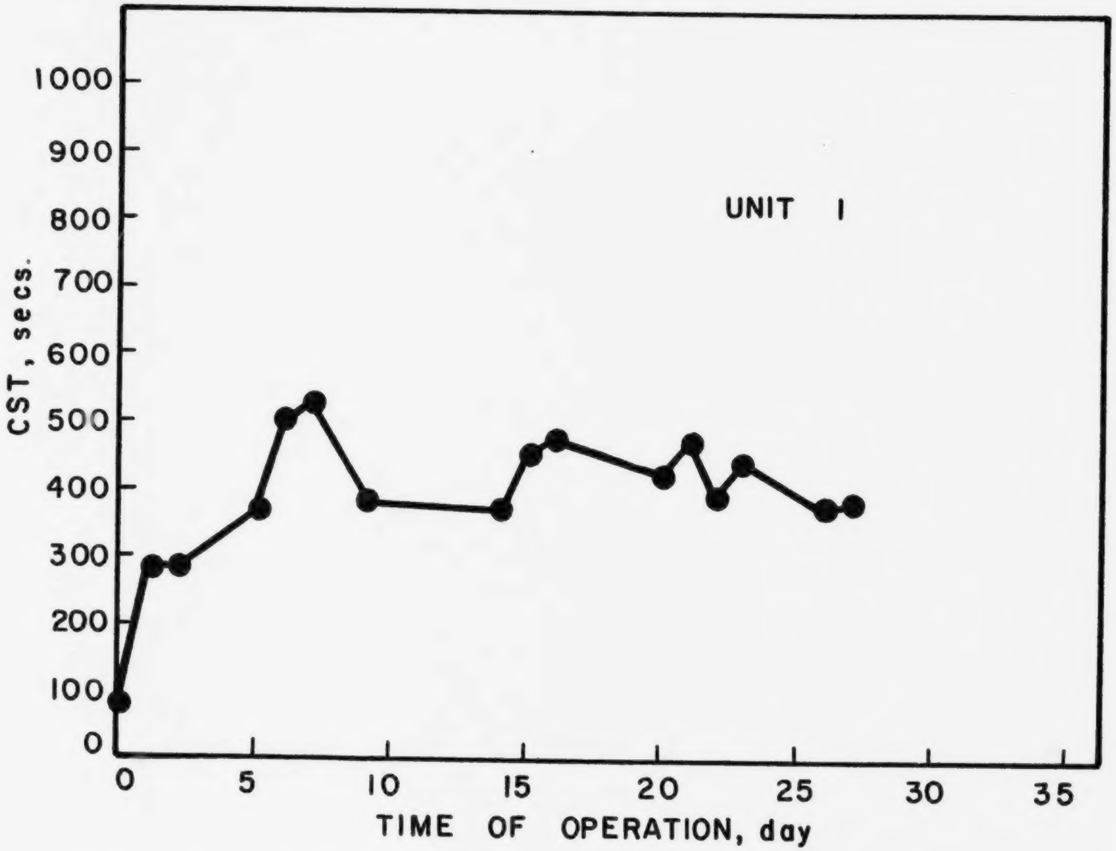

Figure C-101. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 7.

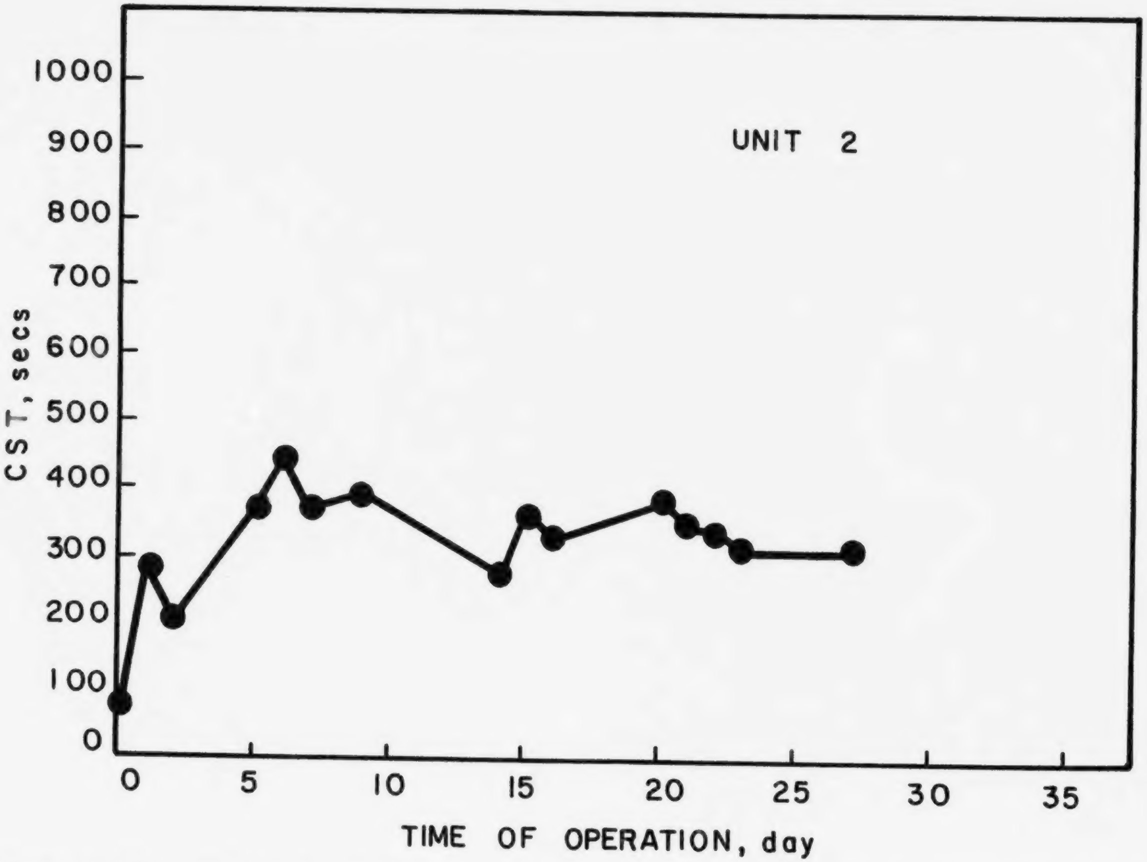

Figure C-102. Change in sludge dewaterability (CST) with time of operation, biodegradability study no. 7.

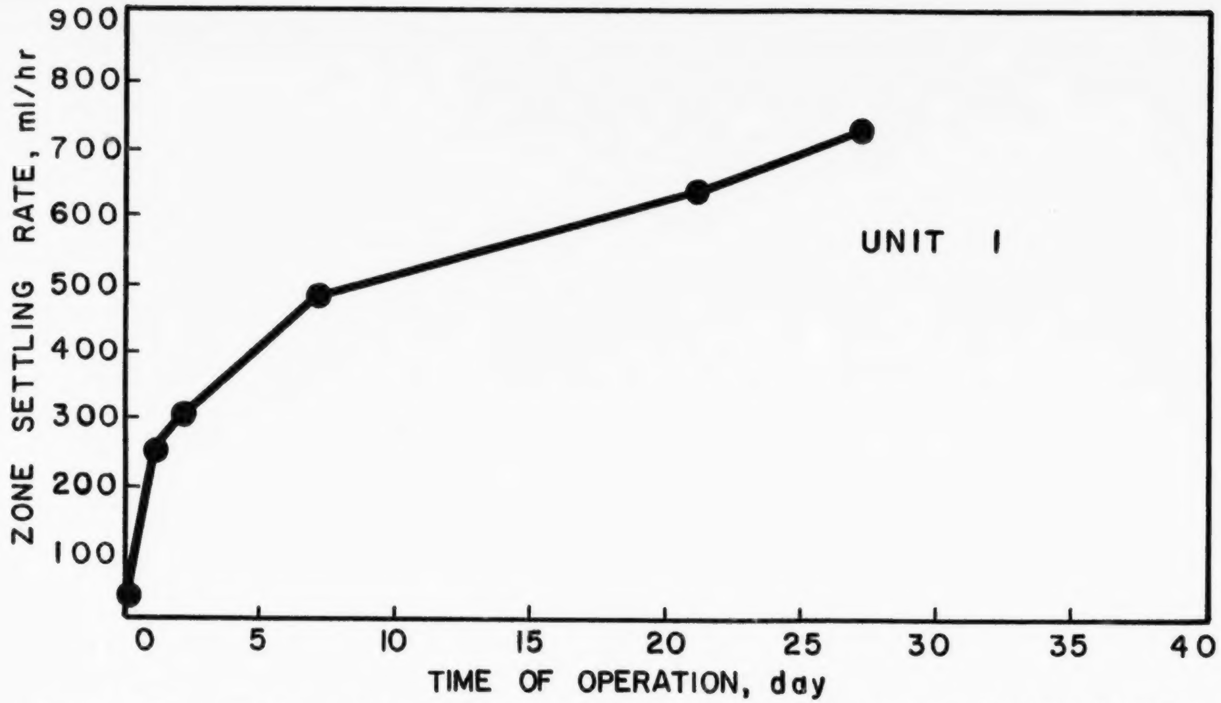

Figure C-103. Zone settling rate (ZSR) as related to time of operation, biodegradability study no. 7.

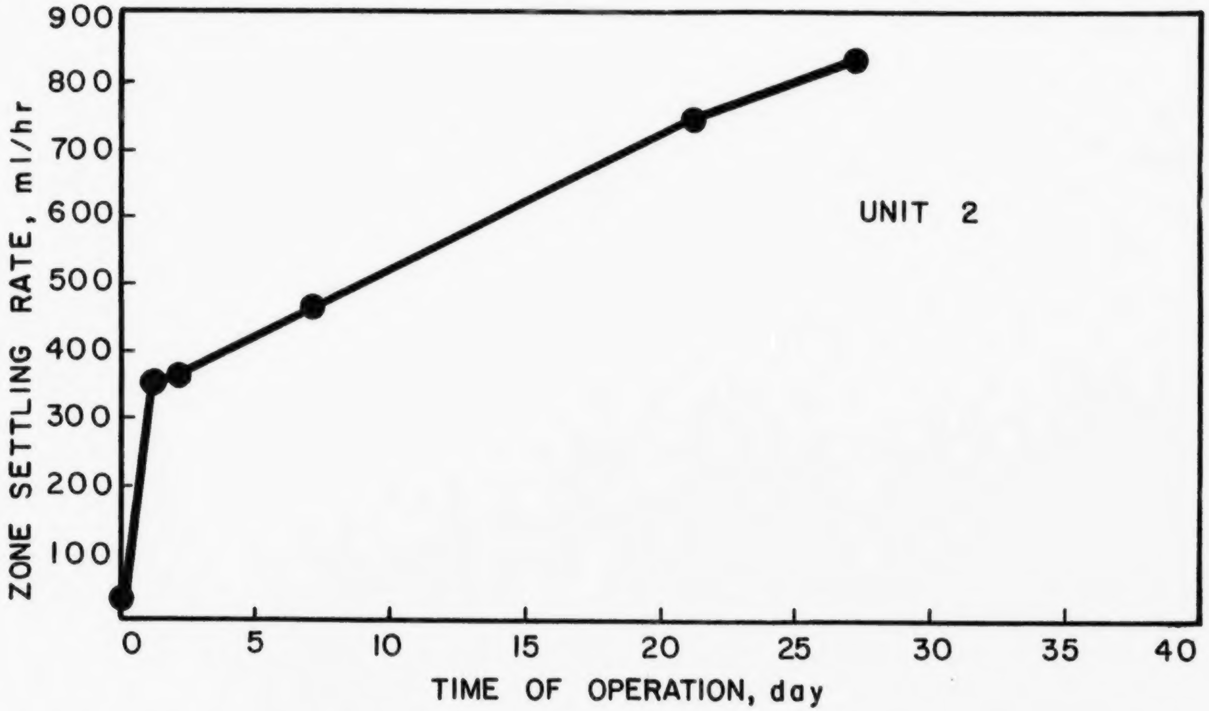

Figure C-104. Zone settling rate (ZSR) as related to time of operation, biodegradability study no. 7.

TABLE D1. PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 3 DAYS (CENTRI-RATOR AERATOR).

| | IDENTIFICATION AND DATES OF STUDY (1978) | | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|---|
| | 4/24-27<br>#1 | 4/27-30<br>#2 | 4/30-5/3<br>#3 | 5/3-6<br>#4 | 5/6-9<br>#5 | 5/9-12<br>#6* | 5/12-15<br>#7* | 5/15-18<br>#8* | 5/18-21<br>#9* | 5/21-24<br>#10 |
| θ Days | 3.03 | 3.07 | 3.05 | 2.73 | 3.25 | 2.83 | 3.11 | 3.64 | 3.21 | 3.27 |
| TCOD Loading Rate** | 21.8 | 21.0 | 18.9 | 20.6 | 18.2 | 19.8 | 18.7 | 14.6 | 17.7 | 16.6 |
| TS Loading Rate** | 20.7 | 18.9 | 17.3 | 18.9 | 17.2 | 19.8 | 19.3 | 14.9 | 17.7 | 17.2 |
| TVS Loading Rate** | 13.5 | 12.4 | 11.3 | 12.2 | 10.5 | 12.1 | 11.8 | 9.22 | 10.9 | 10.4 |
| TCOD Removal Rate** | 5.26 | 2.86 | 3.11 | 1.80 | 1.85 | 4.57 | 5.19 | 3.52 | 4.82 | 3.06 |
| TS Removal Rate** | 3.53 | 1.67 | 1.29 | 1.49 | 1.89 | 3.38 | 4.23 | 1.48 | 1.87 | 2.07 |
| TVS Removal Rate** | 3.87 | 3.04 | 2.27 | 1.85 | 1.64 | 2.84 | 3.78 | 2.09 | 2.47 | 1.98 |
| TCOD Treatment Efficiency (%) | 24.2 | 13.6 | 16.5 | 8.70 | 10.2 | 23.0 | 27.7 | 24.2 | 27.2 | 18.4 |
| TS Treatment Efficiency (%) | 17.1 | 8.80 | 7.50 | 7.90 | 10.9 | 17.1 | 21.9 | 10.0 | 10.6 | 12.0 |
| TVS Treatment Efficiency (%) | 28.7 | 24.6 | 20.1 | 15.1 | 15.5 | 23.5 | 32.0 | 22.6 | 22.5 | 18.9 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | 2744 | - - | 2268 | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | 2107 | - - | 2023 | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | 256 | - - | 188 | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | 485 | - - | 630 | - - | - - | - - |
| pH Feed | 5.6 | 5.7 | 5.8 | 5.9 | 5.7 | 5.8 | 5.7 | - - | 5.6 | 5.6 |
| pH Effluent | 6.7 | 7.0 | 7.1 | 5.7 | 6.0 | 6.8 | 6.8 | - - | 6.9 | 6.7 |
| Temp. Reactor (°C) | 54.9 | 53.6 | 46.6 | 36.0 | 42.0 | 46.9 | 50.4 | 52.9 | 49.8 | 45.8 |
| Temp. Ambient (°C) | 8.1 | 9.5 | 5.2 | 7.3 | 10.7 | 14.1 | 14.6 | 12.6 | 16.5 | 14.6 |
| Temp. Feed (°C) | 9.1 | 9.6 | 10.0 | 9.9 | 10.8 | 10.8 | 11.8 | 12.1 | 12.6 | 13.2 |
| Capillary Suction Time Feed (secs) | 2232 | 1979 | 1813 | 1236 | 830 | 531 | 759 | 332 | 377 | 411 |
| Capillary Suction Time Effluent (secs) | 1682 | 2576 | 1696 | 1342 | 1172 | 1084 | 1018 | 1223 | 1053 | 812 |
| Airflow (ℓ/sec) | 29.8 | 30.4 | 27.0 | 26.0 | 25.6 | 25.9 | 26.0 | 25.5 | 26.5 | 25.9 |
| Oxygen Transfer Efficiency (G.C.)(%) | 23.1 | - - | 22.8 | - - | 18.2 | - - | 23.1 | 22.8 | - - | 21.9 |
| Oxygen Transfer Rate (G.C.)(kg/kW-hr) | 2.05 | - - | 1.86 | - - | 1.25 | - - | 1.98 | 1.90 | - - | 1.77 |
| Oxygen Transfer Efficiency (COD Basis)(%) | 22.5 | 12.3 | 15.1 | 9.13 | 9.73 | 19.5 | 21.6 | 14.9 | 19.4 | 12.9 |
| Oxygen Transfer Rate (COD Basis)(kg/kW-hr) | 1.98 | 1.00 | 1.14 | 0.63 | 0.64 | 1.67 | 1.86 | 1.24 | 1.71 | 1.04 |
| Duration of Test (day) | 2.54 | 2.60 | 2.98 | 2.92 | 3.00 | 2.83 | 2.98 | 2.85 | 2.92 | 2.94 |

*Steady state averaging points.

**$kg/m^3$-day.

TABLE D2. PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 3 DAYS (MIDLAND-FRINGS AERATOR).

| | IDENTIFICATION AND DATES OF STUDY (1978) | | |
|---|---|---|---|
| | 7/26-29 #11 | 7/29-8/1 #12* | 7/24-26 #13 |
| $\theta_H$ (days) | 2.87 | 3.49 | 4.66 |
| TCOD Loading Rate** | 13.6 | 14.6 | 12.5 |
| TS Loading Rate** | 13.5 | 15.6 | 8.56 |
| TVS Loading Rate** | 8.78 | 9.75 | 5.77 |
| TCOD Removal Rate** | 4.53 | 6.53 | 5.14 |
| TS Removal Rate** | 4.64 | 8.30 | 3.13 |
| TVS Removal Rate** | 2.80 | 4.89 | 1.98 |
| TCOD Treatment Efficiency (%) | 33.0 | 44.6 | 44.2 |
| TS Treatment Efficiency (%) | 31.9 | 50.2 | 36.6 |
| TVS Treatment Efficiency (%) | 34.3 | 53.3 | 34.4 |
| TKN Feed (mg/ℓ) | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - |
| pH Feed | - - | 5.4 | - - |
| pH Effluent | - - | 6.80 | - - |
| Temp. Reactor (°C) | 62.0 | 61.8 | 59.9 |
| Temp. Ambient (°C) | 21.0 | 17.7 | 23.3 |
| Temp. Feed (°C) | 20.5 | 20.0 | 21.5 |
| Capillary Suction Time Feed (secs) | 241 | 75 | - - |
| Capillary Suction Time Effluent (secs) | 628 | 553 | 698 |
| Airflow (ℓ/sec) | 39.6 | 39.6 | 40.0 |
| Oxygen Transfer Efficiency (COD Basis) (%) | 15.0 | 22.1 | 17.5 |
| Oxygen Transfer Rate (COD Basis) (kg/kW-hr) | 1.28 | 1.80 | 1.46 |
| Duration of Test (days) | 2.83 | 3.06 | 3.96 |

*Used for steady state averaging.

**kg/m$^3$-day.

## TABLE D3. PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 5 DAYS (CENTRI-RATOR AERATOR).

| | IDENTIFICATION AND DATES OF STUDY (1978) | | | | | | |
|---|---|---|---|---|---|---|---|
| | #14<br>2/20-25 | #15<br>2/25-3/2 | #16<br>3/2-7 | #17*<br>3/7-11 | #18*<br>3/11-15 | #19*<br>3/15-20 | #20*<br>3/20-24 |
| θ Days | 5.71 | 5.55 | 4.60 | 4.67 | 4.39 | 4.77 | 4.41 |
| TCOD Loading Rate** | 9.40 | 9.86 | 11.1 | 11.1 | 11.9 | 12.0 | 13.3 |
| TS Loading Rate** | 9.75 | 9.39 | 10.3 | 10.8 | 11.0 | 12.4 | 13.9 |
| TVS Loading Rate** | 6.08 | 5.94 | 6.60 | 6.93 | 6.98 | 7.55 | 8.44 |
| TCOD Removal Rate** | 3.72 | 2.68 | 2.77 | 3.23 | 3.26 | 4.32 | 4.11 |
| TS Removal Rate** | 3.25 | 0.64 | 0.38 | 1.99 | 2.12 | 3.31 | 2.88 |
| TVS Removal Rate** | 2.62 | 1.29 | 1.20 | 2.05 | 2.01 | 2.71 | 2.64 |
| TCOD Treatment Efficiency (%) | 39.6 | 27.2 | 25.0 | 29.1 | 27.3 | 36.1 | 30.8 |
| TS Treatment Efficiency (%) | 33.4 | 6.87 | 3.70 | 18.4 | 19.4 | 26.7 | 20.7 |
| TVS Treatment Efficiency (%) | 43.0 | 21.7 | 18.2 | 29.5 | 28.7 | 35.9 | 31.2 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | - - | - - | - - |
| pH Feed | - - | - - | 5.7 | 5.7 | 5.8 | - - | 5.9 |
| pH Effluent | - - | - - | 7.1 | 6.9 | 6.8 | 6.9 | 7.1 |
| Temp. Reactor (°C) | 37.6 | 42.5 | 46.7 | 49.0 | 48.0 | 46.9 | 47.9 |
| Temp. Ambient (°C) | -7.1 | -6.3 | -8.3 | -3.4 | 3.0 | 0.3 | 4.0 |
| Temp. Feed (°C) | 5.2 | 5.8 | 5.3 | 5.5 | 6.5 | 6.4 | 6.6 |
| Capillary Suction Time Feed (secs) | - - | 411 | 426 | 257 | 455 | 229 | 690 |
| Capillary Suction Time Effluent (secs) | - - | 592 | 703 | 1012 | 866 | 879 | 740 |
| Airflow (ℓ/sec) | 24.0 | 25.0 | 25.0 | 25.0 | 25.0 | 26.0 | 26.0 |
| Oxygen Transfer Efficiency (G.C.) (%) | - - | - - | - - | - - | - - | - - | - - |
| Oxygen Transfer Rate (G.C.) (kg/kW-hr) | - - | - - | - - | - - | - - | - - | - - |
| Oxygen Transfer Efficiency (COD Basis) (%) | 19.5 | 13.5 | 13.9 | 16.2 | 16.4 | 20.9 | 19.8 |
| Oxygen Transfer Rate (COD Basis) (kg/kW-hr) | 1.19 | 0.85 | 0.89 | 1.02 | 1.05 | 1.47 | 1.32 |
| Duration of Test (day) | 4.90 | 4.79 | 4.94 | 4.17 | 4.23 | 4.56 | 3.92 |

*Steady state averaging.

**kg/m$^3$-day

TABLE D4. PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 7 DAYS (CENTRI-RATOR AERATOR PRIOR TO ADDITION OF REACTOR RADIUS).

| | IDENTIFICATION AND DATES OF STUDY (1977) | | | | | |
|---|---|---|---|---|---|---|
| | #21<br>7/25-8/1 | #22<br>8/1-8 | #23<br>8/8-15 | #24*<br>8/15-22 | #25*<br>8/22-29 | #26*<br>8/29-9/5 |
| $\theta$ Days | 7.28 | 7.14 | 8.02 | 7.74 | 7.20 | 8.27 |
| TCOD Loading Rate** | 3.37 | 3.94 | 4.29 | 5.05 | 5.30 | 5.18 |
| TS Loading Rate** | 4.10 | 4.31 | 4.67 | 4.41 | 4.72 | 4.06 |
| TVS Loading Rate** | 2.64 | 2.82 | 2.96 | 2.73 | 2.92 | 2.68 |
| TCOD Removal Rate** | 0.87 | 1.67 | 1.99 | 2.28 | 2.37 | 2.41 |
| TS Removal Rate** | 0.00 | 1.38 | 1.68 | 1.09 | 1.42 | 0.69 |
| TVS Removal Rate** | 0.41 | 1.16 | 1.25 | 0.84 | 1.11 | 0.90 |
| TCOD Treatment Efficiency (%) | 26.0 | 42.4 | 46.5 | 45.1 | 44.8 | 46.5 |
| TS Treatment Efficiency (%) | 0.00 | 32.1 | 35.9 | 24.7 | 30.2 | 17.0 |
| TVS Treatment Efficiency (%) | 15.6 | 41.1 | 42.1 | 30.9 | 37.9 | 33.8 |
| TKN Feed (mg/ℓ) | - - | - - | - - | 1279 | 1139 | 1209 |
| TKN Effluent (mg/ℓ) | - - | - - | - - | 968* | 1070 | 1420 |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | 103* | 181 | 135 |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | 235* | 280 | 417 |
| pH Feed | 5.5 | 5.2 | 5.4 | 5.5 | - - | 5.5 |
| pH Effluent | 7.1 | 7.5 | 7.6 | 7.6 | 8.0 | 7.5 |
| Temp. Reactor (°C) | 46.9 | 48.7 | 48.5 | 47.1 | 45.2 | 47.8 |
| Temp. Ambient (°C) | 19.0 | 22.7 | 22.6 | 16.6 | 18.8 | 21.7 |
| Temp. Feed (°C) | 21.5 | 22.0 | 22.4 | 21.1 | 21.4 | 22.3 |
| Capillary Suction Time Feed (secs) | - - | - - | - - | - - | - - | - - |
| Capillary Suction Time Effluent (secs) | - - | - - | - - | - - | - - | - - |
| Airflow (ℓ/sec) | 28.5 | 27.0 | 27.4 | 24.3 | 25.4 | 25.2 |
| Oxygen Transfer Efficiency (G.C.) (%) | 6.1 | 10.5 | 7.8 | 11.6 | 7.3 | 7.1 |
| Oxygen Transfer Rate (G.C.) (kg/kW-hr) | 0.31 | 0.57 | 0.50 | 0.74 | 0.51 | 0.52 |
| Oxygen Transfer Efficiency (COD Basis) (%) | 3.2 | 9.1 | 10.6 | 13.7 | 13.7 | 14.0 |
| Oxygen Transfer Rate (COD Basis) (kg/kW-hr) | 0.31 | 0.60 | 0.79 | 0.85 | 0.87 | 0.88 |
| Duration of Test (day) | 7.08 | 6.88 | 7.00 | 6.88 | 7.00 | 7.00 |

*Steady state averaging

**kg/m$^3$-day

TABLE D5. PERFORMANCE DATA AS COLLECTED FROM THE FULL SCALE AEROBIC DIGESTER AT A DESIGN HRT OF 10 DAYS (MIDLAND-FRINGS AERATOR).

| | IDENTIFICATION AND DATES OF STUDY (1978) | | | |
|---|---|---|---|---|
| | #27<br>10/16-26 | #28<br>10/26-11/5 | #29*<br>11/5-15 | #30<br>11/15-25 |
| θ Days | 11.0 | 10.2 | 11.0 | 10.2 |
| TCOD Loading Rate** | 3.40 | 3.76 | 4.36 | 4.23 |
| TS Loading Rate** | 3.32 | 3.56 | 3.49 | 3.50 |
| TVS Loading Rate** | 2.41 | 2.58 | 2.53 | 2.60 |
| TCOD Removal Rate** | 1.02 | 1.52 | 2.51 | 1.77 |
| TS Removal Rate** | 1.10 | 0.82 | 1.43 | 0.93 |
| TVS Removal Rate** | 0.93 | 0.79 | 1.25 | 0.86 |
| TCOD Treatment Efficiency (%) | 30.1 | 39.2 | 57.5 | 41.9 |
| TS Treatment Efficiency (%) | 33.2 | 23.0 | 41.1 | 26.7 |
| TVS Treatment Efficiency (%) | 38.8 | 30.7 | 49.4 | 33.2 |
| TKN Feed (mg/ℓ) | - - | - - | 1435 | - - |
| TKN Effluent (mg/ℓ) | - - | - - | 1323 | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | 105 | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | 650 | - - |
| pH Feed | 5.5 | 5.4 | 5.7 | 5.6 |
| pH Effluent | 6.8 | 6.7 | 7.1 | 7.1 |
| Temp. Reactor (°C) | 54.6 | 60.9 | 62.3 | 65.0 |
| Temp. Ambient (°C) | 7.1 | 6.4 | 6.8 | 2.7 |
| Temp. Feed (°C) | 13.6 | 12.7 | 11.9 | 10.7 |
| Capillary Suction Time Feed (secs) | 500 | 1.0 | 44 | - - |
| Capillary Suction Time Effluent (secs) | 573 | 100 | 145 | - - |
| Airflow (ℓ/sec) | 40.0 | 40.0 | 40.0 | 40.0 |
| Oxygen Transfer Efficiency (G.C.) (%) | - - | - - | - - | - - |
| Oxygen Transfer Rate (G.C.) (kg/kW-hr) | - - | - - | - - | - - |
| Oxygen Transfer Efficiency (COD Basis) (%) | 3.2 | 4.8 | 7.9 | 5.7 |
| Oxygen Transfer Rate (COD Basis) (kg/kW-hr) | 0.29 | 0.40 | 0.66 | 0.45 |
| Duration of Test (day) | 9.88 | 9.95 | 9.94 | 10.0 |

*Steady state averaging

**kg/$m^3$-day

## TABLE D6. PERFORMANCE DATA AS COLLECTED FROM A CONTINUOUSLY-FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 3 DAYS.

| | IDENTIFICATION* | | | | |
|---|---|---|---|---|---|
| | #1** | #2** | #3** | #4** | #5 |
| θ Days | 2.85 | 2.86 | 2.86 | 2.88 | 2.96 |
| TCOD Loading Rate (kg/m$^3$-day) | 13.1 | 12.8 | 13.1 | 13.8 | 10.6 |
| TS Loading Rate (kg/m$^3$-day) | 11.9 | 12.3 | 13.4 | 12.9 | 11.2 |
| TVS Loading Rate (kg/m$^3$-day) | 7.33 | 7.64 | 8.53 | 7.93 | 7.06 |
| TCOD Removal Rate (kg/m$^3$-day) | 4.12 | 4.14 | 5.61 | 6.65 | 2.18 |
| TS Removal Rate (kg/m$^3$-day) | 1.07 | 0.89 | 1.99 | 2.19 | 0.62 |
| TVS Removal Rate (kg/m$^3$-day) | 1.99 | 1.55 | 3.33 | 3.50 | 1.29 |
| TCOD Treatment Efficiency (%) | 31.6 | 32.3 | 42.8 | 48.3 | 20.6 |
| TS Treatment Efficiency (%) | 22.1 | 17.7 | 36.2 | 41.4 | 13.1 |
| TVS Treatment Efficiency (%) | 27.2 | 20.6 | 39.1 | 44.0 | 18.3 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| pH Feed | 6.5 | 6.9 | 5.5 | 7.0 | 6.7 |
| pH Effluent | 7.1 | 7.3 | 6.3 | 7.1 | 6.9 |
| Temp. Reactor (°C) | 55.0 | 55.0 | 55.0 | 56.0 | 54.0 |
| Temp. Ambient (°C) | 15.0 | 15.0 | 15.0 | 15.0 | 15.0 |
| Temp. Water Bath (°C) | 74.0 | 63.0 | 65.0 | 64.0 | 62.0 |
| Capillary Suction Time Feed (secs) | - - | - - | - - | - - | - - |
| Capillary Suction Time Effluent (secs) | - - | - - | - - | - - | - - |
| Airflow (ℓ/sec) | 2.0 | 2.0 | 2.0 | 2.0 | 2.0 |
| Average Reactor Dissolved Oxygen (mg/ℓ) | 0.3 | 1.1 | 1.7 | 1.0 | 0.1 |

[*] Study covered a period from July 8, 1977 to September 3, 1977. The HRT is equal to the duration of the test in all cases.

[**] Steady state averaging points

## TABLE D7. PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 5 DAYS, FIRST RUN.

| | IDENTIFICATION* | | | | |
|---|---|---|---|---|---|
| | #6 | #7** | #8** | #9** | #10 |
| # Days | 5.0 | 5.0 | 5.0 | 5.0 | 5.0 |
| TCOD Loading Rate (kg/m$^3$-day) | 8.78 | 10.8 | 11.8 | 10.7 | 9.92 |
| TS Loading Rate (kg/m$^3$-day) | 12.0 | 9.41 | 10.8 | 10.3 | 9.45 |
| TVS Loading Rate (kg/m$^3$-day) | 6.18 | 6.00 | 6.63 | 6.16 | 6.26 |
| TCOD Removal Rate (kg/m$^3$-day) | 3.77 | 4.20 | 4.41 | 3.81 | 3.65 |
| TS Removal Rate (kg/m$^3$-day) | 2.35 | 1.91 | 2.85 | 2.24 | 2.75 |
| TVS Removal Rate (kg/m$^3$-day) | 2.31 | 1.61 | 1.87 | 1.53 | 2.36 |
| TCOD Treatment Efficiency (%) | 42.9 | 38.9 | 37.3 | 35.6 | 36.8 |
| TS Treatment Efficiency (%) | 51.4 | 20.3 | 26.4 | 21.9 | 29.1 |
| TVS Treatment Efficiency (%) | 40.8 | 26.7 | 28.2 | 24.8 | 37.7 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| pH Feed | 5.9 | 5.8 | 5.8 | 5.8 | 5.8 |
| pH Effluent | 6.8 | 6.8 | 6.7 | 6.7 | 6.5 |
| Temp. Reactor (°C) | 49.0 | 42.0 | 51.0 | 50.1 | 50,0 |
| Temp. Ambient (°C) | 15.0 | 15.0 | 15.0 | 15.0 | 15.0 |
| Temperature (°C) | 65.0 | 50.0 | 70.0 | 70.0 | 70.0 |
| Capillary Suction Time Feed (secs) | - - | - - | - - | - - | - - |
| Capillary Suction Time Effluent (secs) | - - | - - | - - | - - | - - |
| Airflow (ℓ/sec) | 2.0 | 2.0 | 2.0 | 2.0 | 2.0 |
| Average Reactor Dissolved Oxygen (mg/ℓ) | - - | - - | - - | 3.7 | 2.6 |

*Study covered a period from October 1, 1977 to December 19, 1977. The HRT is equal to the duration of the test in all cases.

**Steady state averaging.

## TABLE D8. PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY-FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 5 DAYS, SECOND RUN.

| | IDENTIFICATION* | | | | | |
|---|---|---|---|---|---|---|
| | #11 | #12** | #13** | #14** | #15 | #16 |
| θ Days | 5.0 | 5.0 | 5.0 | 5.0 | 5.0 | 5.0 |
| TCOD Loading Rate (kg/m$^3$-day) | 11.5 | 11.5 | 12.4 | 12.4 | 12.5 | 12.6 |
| TS Loading Rate (kg/m$^3$-day) | 12.2 | 12.2 | 14.2 | 14.2 | 11.7 | 10.5 |
| TVS Loading Rate (kg/m$^3$-day) | 7.01 | 7.01 | 8.29 | 8.29 | 6.97 | 6.64 |
| TCOD Removal Rate (kg/m$^3$-day) | 3.38 | 3.97 | 4.74 | 3.44 | 3.01 | 4.55 |
| TS Removal Rate (kg/m$^3$-day) | 2.41 | 2.39 | 5.09 | 3.66 | 1.91 | 1.56 |
| TVS Removal Rate (kg/m$^3$-day) | 1.84 | 1.93 | 3.46 | 2.56 | 1.37 | 1.39 |
| TCOD Treatment Efficiency (%) | 29.5 | 34.6 | 38.3 | 27.8 | 24.2 | 36.2 |
| TS Treatment Efficiency (%) | 19.8 | 19.7 | 36.0 | 25.9 | 16.3 | 14.9 |
| TVS Treatment Efficiency (%) | 26.2 | 27.5 | 41.7 | 30.9 | 19.7 | 20.9 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | 3010 | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | 2156 | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | 42 | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | 62 | - - |
| pH Feed | 6.0 | 6.0 | 6.0 | 6.0 | 5.7 | 5.7 |
| pH Effluent | 7.6 | 7.5 | 7.5 | 7.9 | 8.0 | 8.1 |
| Temp. Reactor (°C) | 50.0 | 51.0 | 50.7 | 52.8 | 51.7 | 50.5 |
| Temp. Ambient (°C) | 15.0 | 15.0 | 15.0 | 15.0 | 15.0 | 15.0 |
| Temp. Feed (°C) | 69.0 | 69.0 | 70.0 | 69.0 | 69.0 | 65.0 |
| Capillary Suction Time Feed (secs) | 964 | 964 | 964 | 964 | 1203 | 517 |
| Capillary Suction Time Effluent (secs) | 1033 | 1235 | 2077 | 4032 | 2326 | 1877 |
| Airflow (ℓ/sec) | 2.8 | 2.8 | 2.8 | 2.8 | 2.8 | 2.8 |
| Average Reactor Dissolved Oxygen (mg/ℓ) | - - | - - | - - | - - | - - | - - |

*Study covered a period from January 25, 1978 to May 12, 1978. The HRT is equal to the duration of the test in all cases.

**Used for steady state averaging

## TABLE D9. PERFORMANCE DATA AS COLLECTED FROM A SEMI-CONTINUOUSLY FED BENCH SCALE DIGESTER AT A DESIGN HRT OF 7 DAYS.

| | IDENTIFICATION* | | | | |
|---|---|---|---|---|---|
| | #17 | #18 | #19 | #20** | #21** |
| $\theta$ Days | 7.0 | 7.0 | 7.0 | 7.0 | 7.0 |
| TCOD Loading Rate (kg/$m^3$-day) | 6.49 | 5.11 | 5.18 | 5.32 | 5.94 |
| TS Loading Rate (kg/$m^3$-day) | 6.50 | 5.31 | 4.88 | 5.44 | 5.75 |
| TVS Loading Rate (kg/$m^3$-day) | 4.10 | 3.23 | 2.82 | 3.56 | 3.85 |
| TCOD Removal Rate (kg/$m^3$-day) | 3.05 | 2.05 | 2.31 | 2.66 | 2.74 |
| TS Removal Rate (kg/$m^3$-day) | 2.83 | 1.78 | 1.24 | 2.07 | 2.13 |
| TVS Removal Rate (kg/$m^3$-day) | 1.78 | 1.18 | 0.84 | 1.84 | 1.74 |
| TCOD Treatment Efficiency (%) | 47.0 | 40.0 | 45.0 | 50.0 | 46.0 |
| TS Treatment Efficiency (%) | 44.0 | 33.0 | 25.0 | 38.0 | 37.0 |
| TVS Treatment Efficiency (%) | 43.0 | 37.0 | 30.0 | 52.0 | 45.0 |
| TKN Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| TKN Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Feed (mg/ℓ) | - - | - - | - - | - - | - - |
| $NH_3$-N Effluent (mg/ℓ) | - - | - - | - - | - - | - - |
| pH Feed | 5.9 | 5.9 | 6.1 | 6.1 | 5.3 |
| pH Effluent | 6.4 | 6.8 | 6.4 | 6.8 | 7.4 |
| Temp. Reactor (°C) | 49.0 | 56.0 | 55.0 | 55.0 | 54.0 |
| Temp. Ambient (°C) | 15.0 | 15.0 | 15.0 | 15.0 | 15.0 |
| Temp. Water Bath (°C) | 55.0 | 65.0 | 66.0 | 66.0 | 65.0 |
| Capillary Suction Time Feed (secs) | - - | - - | - - | - - | - - |
| Capillary Suction Time Effluent (secs) | - - | - - | - - | - - | - - |
| Airflow (ℓ/sec) | 1.9 | 1.9 | 1.9 | 1.9 | 1.9 |
| Average Reactor Dissolved Oxygen (mg/ℓ) | 3.4 | 3.6 | 3.8 | 3.1 | 1.7 |

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[*] Study covered a period from August 11, 1977 to September 19, 1977. The HRT is equal to the duration of the test in all cases.

[**] Used for steady state averaging.

TABLE D-10. SOLIDS DATA AS COLLECTED FROM THE FULL SCALE DIGESTER AT ALL CONDITIONS TESTED

| Study I.D. | Hydraulic Retention Period (Day) | Influent Solids Total Solids gm/ℓ | Influent Solids Total Volatile Solids gm/ℓ | Effluent Solids Total Solids gm/ℓ | Effluent Solids Total Volatile Solids gm/ℓ |
|---|---|---|---|---|---|
| 1 | 3.03 | 62.6 | 40.9 | 51.9 | 29.1 |
| 2 | 3.07 | 58.2 | 37.9 | 53.0 | 28.6 |
| 3 | 3.05 | 52.6 | 34.4 | 48.7 | 27.5 |
| 4 | 2.73 | 51.7 | 33.5 | 47.7 | 28.4 |
| 5 | 3.25 | 55.9 | 34.3 | 49.8 | 28.9 |
| 6 | 2.83 | 56.1 | 34.3 | 46.6 | 26.3 |
| 7 | 3.11 | 60.1 | 36.7 | 47.0 | 25.0 |
| 8 | 2.64 | 54.1 | 33.6 | 48.7 | 26.0 |
| 9 | 3.21 | 56.9 | 35.3 | 50.9 | 27.3 |
| 10 | 3.27 | 56.3 | 34.1 | 49.5 | 27.7 |
| 11 | 2.87 | 38.3 | 25.2 | 25.5 | 17.2 |
| 12 | 3.49 | 54.4 | 34.1 | 25.4 | 17.0 |
| 13 | 4.66 | 39.9 | 26.9 | 25.3 | 17.6 |
| 14 | 5.71 | 55.7 | 34.8 | 37.1 | 19.8 |
| 15 | 5.55 | 52.1 | 33.0 | 48.5 | 25.8 |
| 16 | 4.60 | 47.2 | 30.4 | 45.5 | 24.9 |
| 17 | 4.67 | 50.4 | 32.3 | 41.2 | 22.8 |
| 18 | 4.39 | 48.1 | 30.6 | 38.8 | 21.8 |
| 19 | 3.77 | 59.1 | 36.0 | 43.4 | 23.1 |
| 20 | 4.41 | 61.2 | 37.2 | 48.7 | 25.6 |
| 21 | 7.28 | 29.9 | 19.2 | 30.6 | 16.2 |
| 22 | 7.14 | 30.8 | 20.1 | 20.9 | 11.9 |
| 23 | 8.02 | 37.4 | 23.8 | 24.0 | 13.8 |
| 24 | 7.74 | 34.1 | 21.1 | 25.7 | 14.6 |
| 25 | 7.20 | 34.0 | 21.0 | 23.7 | 13.0 |
| 26 | 8.27 | 33.6 | 22.2 | 22.9 | 14.7 |
| 27 | 11.0 | 36.6 | 26.5 | 24.4 | 16.2 |
| 28 | 10.2 | 36.5 | 26.5 | 28.1 | 18.3 |
| 29 | 11.0 | 38.2 | 27.7 | 22.5 | 14.0 |
| 30 | 10.2 | 35.8 | 26.6 | 26.2 | 17.8 |

TABLE D-11. SOLIDS DATA AS COLLECTED FROM THE BENCH SCALE DIGESTER AT ALL CONDITIONS TESTED

| | | Influent Solids | | Effluent Solids | |
|---|---|---|---|---|---|
| Study I.D. | Hydraulic Retention Period (Day) | Total Solids gm/ℓ | Total Volatile Solids gm/ℓ | Total Solids gm/ℓ | Total Volatile Solids gm/ℓ |
| 1 | 2.85 | 33.8 | 20.9 | 26.3 | 15.2 |
| 2 | 2.86 | 35.1 | 21.8 | 28.9 | 17.4 |
| 3 | 2.86 | 38.4 | 24.4 | 24.5 | 14.9 |
| 4 | 2.88 | 37.0 | 22.9 | 21.7 | 12.8 |
| 5 | 2.96 | 33.2 | 20.9 | 28.9 | 17.1 |
| 6 | 5.0 | 60.1 | 28.4 | 29.2 | 16.8 |
| 7 | 5.0 | 47.1 | 30.0 | 37.5 | 22.0 |
| 8 | 5.0 | 53.9 | 33.1 | 39.7 | 23.8 |
| 9 | 5.0 | 53.4 | 30.9 | 40.1 | 23.1 |
| 10 | 5.0 | 47.2 | 31.2 | 33.5 | 19.5 |
| 11 | 5.0 | 60.8 | 35.1 | 48.8 | 25.9 |
| 12 | 5.0 | 60.8 | 35.1 | 48.9 | 25.4 |
| 13 | 5.0 | 70.8 | 41.4 | 45.3 | 24.1 |
| 14 | 5.0 | 70.8 | 41.4 | 52.5 | 28.6 |
| 15 | 5.0 | 58.6 | 34.8 | 49.0 | 27.9 |
| 16 | 5.0 | 52.6 | 33.2 | 44.7 | 26.2 |
| 17 | 7.0 | 45.5 | 28.6 | 25.7 | 16.2 |
| 18 | 7.0 | 37.1 | 22.6 | 24.7 | 14.3 |
| 19 | 7.0 | 34.2 | 19.7 | 25.5 | 13.8 |
| 20 | 7.0 | 38.1 | 24.9 | 23.7 | 12.1 |
| 21 | 7.0 | 40.3 | 26.9 | 25.3 | 14.7 |

TABLE D-12. DISSOLVED OXYGEN LEVELS IN THE FULL SCALE DIGESTER FOR ALL CONDITIONS TESTED

| Study I.D. & Dates of Analysis | Hydraulic Retention Period (Day) | Dissolved Oxygen Levels* at Different Reactor Depths** | | | | | | | | Reactor Temp (°C) |
|---|---|---|---|---|---|---|---|---|---|---|
| | | Reactor Top 0 | 0.305 | 0.610 | 0.914 | 1.220 | 1.525 | 1.830 | Reactor Bottom 2.135 | |
| 1. 4/24/78 (C)† | 3.03 | - | 0.60 | 0.65 | 0.70 | 0.75 | 0.75 | 0.75 | 0.75 | 55 |
| 2. 4/27/78 (C) | 3.07 | - | 0.30 | 0.10 | 0.07 | 0.10 | 0.13 | 0.15 | 0.13 | 55 |
| 3. 5/03/78 (C) | 3.05 | - | 0.50 | 0.15 | 0.10 | 0.10 | 0.12 | 0.15 | 0.13 | 50 |
| 5. 5/08/78 (C) | 3.25 | - | 0.40 | 0.50 | 0.05 | 0.04 | 0.05 | 0.10 | 0.05 | 40 |
| 6. 5/12/78 (C) | 2.83 | - | 2.50 | 1.50 | 0.75 | 0.05 | 0.05 | 0.08 | 0.10 | 47 |
| 7. 5/15/78 (C) | 3.11 | - | 0.10 | 0.15 | 0.05 | 0.03 | 0.03 | 0.03 | 0.04 | 52 |
| 8. 5/18/78 (C) | 3.64 | - | 1.50 | 0.75 | 0.10 | 0.05 | 0.04 | 0.04 | 0.04 | 55 |
| 10. 5/24/78 (C) | 3.27 | - | 1.26 | 0.30 | 0.10 | 0.10 | 0.10 | 0.10 | 0.10 | 45 |
| 11. 7/28/78 (C) | 3.87 | - | 0.20 | 0.10 | 0.10 | 0.15 | 0.10 | 0.10 | 0.05 | 63 |
| 13. 7/24/78 (C) | 4.66 | - | 2.00 | 2.00 | 1.00 | 1.00 | 1.00 | 1.00 | 0.50 | 61 |
| 14. 2/25/78 (C) | 5.71 | 3.80 | 1.20 | 0.50 | 0.20 | 0.70 | 0.60 | 0.80 | 0.70 | 38 |
| 15. 3/02/78 (C) | 5.55 | - | 3.00 | 1.30 | 2.00 | 0.20 | 0.15 | 0.15 | 0.15 | 45 |
| 16. 3/07/78 (C) | 4.60 | 2.50 | 2.50 | 2.30 | 0.35 | 0.30 | 0.65 | 0.65 | 0.65 | 48 |
| 17. 3/09/78 (C) | 4.67 | 1.60 | 1.60 | 1.40 | 0.05 | 0.05 | 0.07 | 0.08 | 0.08 | 49 |
| 18. 3/13/78 (C) | 4.39 | 2.00 | 2.80 | 2.50 | 0.08 | 0.07 | 0.07 | 0.07 | 0.08 | 49 |
| 19. 3/16/78 (C) | 4.77 | 2.60 | 3.00 | 3.00 | 0.13 | 0.13 | 0.13 | 0.14 | 0.14 | 48 |
| 20. 3/24/78 (C) | 4.41 | 1.70 | 1.80 | 1.10 | 0.15 | 0.10 | 0.25 | 0.23 | 0.28 | 48 |
| 21. 7/29/77 (M)† | 7.28 | - | - | - | - | - | - | - | 0.40 | 48 |
| 22. 8/01/77 (M) | 7.14 | - | - | - | 6.40 | - | - | 0.40 | - | 49 |
| 23. 8/08 77 (M) | 8.02 | - | - | - | 6.00 | - | 0.50 | - | 0.40 | 49 |
| 24. 8/16/77 (M) | 7.74 | - | - | - | 6.10 | - | 0.2 | 0.20 | - | 48 |
| 25. 8/23/77 (M) | 7.20 | - | - | - | 5.40 | - | 0.50 | - | 0.20 | 47 |
| 26. 8/30/77 (M) | 8.27 | 4.70 | - | 4.30 | 1.50 | 0.30 | 0.30 | 0.30 | 0.30 | 47 |
| 29. 11/15/78 (M) | 11.0 | - | 2.50 | 1.50 | 1.00 | 1.50 | 2.50 | 0.50 | 0.15 | 63 |

*mg/ℓ. **m.

†(C) = centri-rator; (M) = Midland-Frings; aerators.